

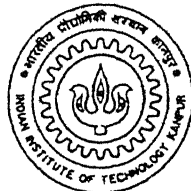
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STUDY OF THE COREX PROCESS OF IRONMAKING THROUGH A THERMO-CHEMICAL MODEL

by

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to the

MATERIALS AND METALLURGICAL ENGINEERING
INDIAN INSTITUTE OF TECHNOLOGY KANPUR
MARCH 1999

STUDY OF THE COREX PROCESS OF IRONMAKING THROUGH A THERMO-CHEMICAL MODEL

A Thesis Submitted

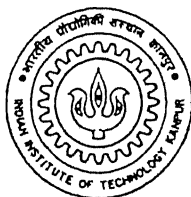
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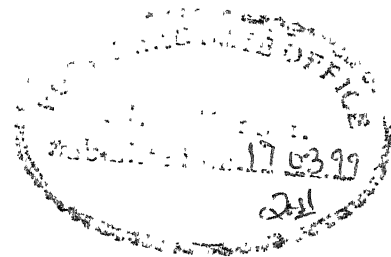
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INDIAN INSTITUTE OF TECHNOLOGY KANPUR
MARCH 1999



CERTIFICATE

It is certified that the work contained in this thesis entitled "*Study of the COREX process of ironmaking through a thermo-chemical model*", by Milan Kumar Barui, has been carried out under my supervision and that this work has not been submitted elsewhere for any degree.


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Dedicated to

My parents

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I wish to express my gratitude to Prof. S .C . KORIA for introducing me in the new and interesting field of ironmaking. . The successful completion of this work has been possible only due to his excellent guidance and sugessions.

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Abstract

COREX process is a two stage smelting reduction process produces hot metal of blast furnace quality. Coal is used instead of coke and there is no need of sintering or, pelletization plant for using fine iron ore. This process has become true alternative of blast furnace in which reduction and smelting are carried out in different reactor. Reduction gas is generated in a fluidized bed by partial combustion of coal and is fed into the reduction shaft furnace for reduction of ore.

A thermo-chemical model on this corex process based on materials and energy balances at steady state has been developed. The commercial scale data of a corex plant are used to verify the model predictions. Model is found to predict the plant data satisfactorily. This model predicts coal consumptions, flux rate, slag rate, oxygen required for coal combustion, medium required for post combustion, export gas amount, gas temperature, volume and it's compositions. It takes into account all of the important variables including degree of metallization, post combustion ratio, heat transfer efficiency, post combustion medium, coal composition. It has been shown that post combustion greatly reduces coal consumption, oxygen required, export gas amount and this post combustion ratio and degree of metallization must be adjusted in such a way that no reforming is required for reduction of iron ore. Further it is shown that heat transfer efficiency and fraction of O_2 in O_2 -air mixture are two important parameter to control the gas temperature which goes up due to post combustion. It is shown that the heat transfer efficiency should be as high as possible to control gas temperature and coal consumption.

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NOMENCLATURE

(All are expressed in per kg mole of product Fe)

$n_{Fe}^i, n_C^i, n_{Si}^i, n_{Mn}^i, n_S^i, n_P^i, n_H^i, n_{Ca}^i, n_O^i, n_N^i, n_{Mg}^i, n_{Al_2O_3}^i$	Kg moles of respective element or, oxide in
$n_{Fe}^o, n_C^o, n_{Si}^o, n_{Mn}^o, n_S^o, n_P^o, n_H^o, n_{Ca}^o, n_O^o, n_N^o, n_{Mg}^o, n_{Al_2O_3}^o$	Kg moles of respective element or, oxide out
$n_C^{coal}, n_H^{coal}, n_S^{coal}, n_N^{coal}, n_O^{coal}$	Kg moles of respective element in coal
$n_{SiO_2}^{ash}, n_{Al_2O_3}^{ash}, n_{CaO}^{ash}, n_{MgO}^{ash}$	Kg moles of respective oxides in ash of coal
$n_{SiO_2}^{ore}, n_{Al_2O_3}^{ore}, n_{FeS}^{ore}, n_{MnO}^{ore}, n_{P_2O_5}^{ore}$	Kg moles of respective compounds in ore
$n_{CaCO_3}^L, n_{MgCO_3}^L, n_{SiO_2}^L, n_{Al_2O_3}^L$	Kg moles of respective compounds in limestone
$n_{CaCO_3}^D, n_{MgCO_3}^D, n_{SiO_2}^D, n_{Al_2O_3}^D$	Kg moles of respective compounds in dolomite
$n_C^{HM}, n_{Si}^{HM}, n_{Mn}^{HM}, n_S^{HM}, n_P^{HM}$	Kg moles of respective element in hot metal
$n_{CaO}^{slag}, n_{CaS}^{slag}, n_{SiO_2}^{slag}, n_{Al_2O_3}^{slag}, n_{MnO}^{slag}, n_{MgO}^{slag}$	Kg moles of respective compounds in slag
$n_{CO}^r, n_{CO_2}^r, n_{H_2}^r, n_{H_2O}^r, n_{N_2}^r$	Kg moles of respective compounds leaving melter-gasifier
$n_{CO}^{rg}, n_{CO_2}^{rg}, n_{H_2}^{rg}, n_{H_2O}^{rg}, n_{N_2}^{rg}$	Kg moles of respective compounds in reduction gas
$n_{CO}^{sg}, n_{CO_2}^{sg}, n_{H_2}^{sg}, n_{H_2O}^{sg}, n_{N_2}^{sg}$	Kg moles of respective compounds leaving in surplus gas
$n_{CO}^{tg}, n_{CO_2}^{tg}, n_{H_2}^{tg}, n_{H_2O}^{tg}, n_{N_2}^{tg}$	Kg moles of respective compounds leaving in the top gas
$n_{Fe}, n_{Fe_{0.95}O}$	Kg moles of Fe and $Fe_{0.95}O$ leaving reduction shaft furnace respectively

$n_C^{tg}, n_H^{tg}, n_N^{tg}$	Kg moles of respective element in the top gas
$n_C^{sg}, n_H^{sg}, n_N^{sg}$	Kg moles of respective element in surplus gas
n_H^M, n_O^M	Kg moles of respective element in moisture of coal
n_O^B	Kg moles of oxygen injected for coal combustion
n_O^{pc}, n_N^{pc}	Kg moles respective element injected for post combustion
n_C^A	Kg moles of carbon reacting with oxygen supplied for coal combustion
$\left(\frac{O}{C}\right)^{sg}$	Kg moles of O per kg moles of C leaving in surplus gas
$\left(\frac{O}{C}\right)^{tg}$	Kg moles of O per kg moles of C leaving in top gas
$\left(\frac{O}{H_2}\right)^{sg}$	Kg moles of O per kg moles of H_2 leaving in surplus gas
$\left(\frac{O}{H_2}\right)^{tg}$	Kg moles of O per kg moles of H_2 leaving in top gas
$\left(\frac{O}{Fe}\right)$	Kg moles of O per kg moles of Fe in iron oxide
M_L	Kg limestone
M_D	Kg dolomite
M_C	Kg coal
M_I	Kg iron ore
ϵ	Kg moles of CO_2 dissociation
ΔH_{HM}	Hot metal heat demand
ΔH_{HM}^s	Sensible heat of constituent in hot metal
ΔH_{HM}^M	Heat of mixing of C, Si, Mn, S, P in hot metal
ΔH_E	Heat absorbed by endothermic reactions in melter gasifier
ΔH_{slag}	Slag heat demand
ΔH_{slag}^s	Sensible heat of slag
ΔH_{slag}^f	Heat of slag formation
ΔH_{gas}	Sensible heat of gas leaving melter gasifier
ΔH_{coal}	Heat demand for drying of coal
ΔH_D	Latent heat of devolatilization of coal

Q_{MG}	Heat losses in melter-gasifier
ΔH_{dem}	Melter-gasifier heat demand
Q_{LS}	Heat losses in reduction shaft furnace
ΔH_{RS}	Reduction shaft furnace heat demand
S_{coal}	Heat supplied by coal combustion
S_{pc}	Heat supplied by post combustion
n_{Si}^a	Kg moles of SiO_2 from ore (joining hot metal ad Si)
$n_{SiO_2}^b$	Kg moles of SiO_2 from ore (joining slag ad SiO_2)
n_{Si}^c	Kg moles of SiO_2 from ash (joining hot metal ad Si)
$n_{SiO_2}^d$	Kg moles of SiO_2 from ash (joining slag ad SiO_2)
$n_{SiO_2}^f$	Kg moles of SiO_2 from flux (joining hot metal ad Si)
$n_{SiO_2}^g$	Kg moles of SiO_2 from flux (joining slag ad SiO_2)
n_S^a	Kg moles of S from ore (joining hot metal as S)
n_{CaS}^b	Kg moles of S from ore (joining slag as CaS)
n_S^c	Kg moles of S from coal (joining hot metal as S)
n_{CaS}^d	Kg moles of S from coal (joining slag as CaS)

$p_C^{coal}, p_H^{coal}, p_S^{coal}, p_N^{coal}, p_O^{coal}$	Percentages of respective element in coal
$p_{SiO_2}^{ash}, p_{Al_2O_3}^{ash}, p_{CaO}^{ash}, p_{MgO}^{ash}$	Percentages of respective compounds in ash
$p_{SiO_2}^{ore}, p_{Al_2O_3}^{ore}, p_{FeS}^{ore}, p_{MnO}^{ore}, p_{P_2O_5}^{ore}$	Percentages of respective compounds in iron ore
$p_{CaCO_3}^L, p_{MgCO_3}^L, p_{SiO_2}^L, p_{Al_2O_3}^L$	Percentages of respective compounds in limestone
$p_{CaCO_3}^D, p_{MgCO_3}^D, p_{SiO_2}^D, p_{Al_2O_3}^D$	Percentages of respective compounds in limestone
$p_C^{HM}, p_{Fe}^{HM}, p_S^{HM}, p_P^{HM}, p_{Mn}^{HM}, p_{Si}^{HM}$	Percentages of respective element in hot metal
$p_{O_2}^{med}, p_{N_2}^{med}$	Percentages of oxygen and nitrogen comes from post combustion medium respectively
p_H^M, p_O^M	Percentages of respective elements in moisture

B_C	Basicity of slag
T_g	Temperature of gas in the melter-gasifier
T_{tg}	Temperature of top gas
T_R	Temperature of reduction gas
X	Degree of metallization
ppc	Post combustion ratio
HTE	Heat transfer efficiency

Chapter 1

Introduction

In the past few decades, the blast furnace process has attained a level of perfection which led to an essential increase in production, improvement of quality and reduction in the consumption of combustibles[2]. Modern blast furnaces using high quality sinter and premium grade coke can produce at a rate as high as $2.9 \text{ t}/\text{m}^3/\text{day}$ [4] with a fuel rate as low as $396 \text{ kg}/\text{t}$ and can work for more than ten years without relining. However, the price is paid for this perfection is the preparation of input materials as coal and ore fines can not be charged there. For this, coal is required to be converted into coke and ore fines into sinter before they are charged in blast furnace. It becomes more and more difficult or, costly to master the environmental problems arising in the production of coke and sinter[2]. Dependence on very high capital investment is also another problem for this because more than 60%[4] of the capital is spent for construction of coke ovens which is used to convert coal to coke and sintering plant which is used to make sinter. Scarcity of coking coal is also another problem. The environmental pollution in steel plants are mainly due to dust, carbon dioxide and pollutants from coke oven effluents such as phenol, cyanide, oil, grease. In blast furnace area, sintering and coke oven plants are primarily responsible for this. So, effort to reduce the coking coal consumption is welcome in steel industry. It is extremely important to modify the blast furnace to a coal based alternatives.

Many advances have been made for reducing the coke consumption. They are increased

blast temperature, oxygen enrichment, fuel injection, high top pressure , use of fluxed burden, proper sizing and size distribution of charge materials, coal injection in the blast furnace and blast humidification[5]. But a blast furnace cannot work without coke, principally because it needs a coke grid in hearth and bosh to support burden and "coke-windows" in the cohesive zone in order to maintain a satisfactory gas permeability in this area[6].

So, there is a need for alternative ironmaking processes which can directly use non-coking coal and run economically on a smaller scale. Direct reduction has been the main solution over the past few decades, during which gas-based processes have proved far more popular than their coal based counterparts. Direct reduction processes employ gas or coal for the reduction of iron ore to sponge iron in solid state. Sponge iron is primarily suitable as a replacement for high quality scrap in electric steelmaking[6]. But with solid state reactions and diffusion in the solid state, the conversion rate remains limited and, on account of the occurrence of process disadvantages such as sticking, it is not possible to utilize the exponential increase of the reaction rate with temperature in spite of a reduction in particle size

A shift to liquid phases opens new possibilities for reduction processes. Not only the increased transport rate due to convection but also the considerable increase in the conversion rate due to the enlargement of the specific phase contact areas in dispersed phases can be advantageously utilized, together with the considerable increase in reaction and diffusion rates due to higher temperature[6]. The quintessence of these considerations is the further development of the ironmaking process in the direction of smelting reduction. The term smelting reduction is used today to designate processes of reduction of iron ore at high temperature to obtain liquid iron using coal (non-coking in general) and oxygen, with a substantial part of the reduction taking place in the liquid oxide phase. The direct use of primary energy instead of electrical energy with a simultaneous gasification of coal requires a combination of final reduction and of melting in or, directly above the iron bath, with the use of the produced gases for pre-reduction.

use of the produced gases for pre-reduction.

The advantages of the smelting reduction processes are as follows :[1][3][7][4][13]

i) The process uses non-coking coal which is undoubtedly cheapest and most widely distributed source of energy. Gases generated by devolatilising and gasifying coal are utilized at different extents for pre-reduction and/or for power generation.

ii) Direct use of ore fines would avoid agglomeration plants.

iii) By using coal directly, it avoids coke oven and in certain cases, pelletisation and sintering plant because of the direct use of ore fines. So, capital cost is very low.

iv) This route of ironmaking is economical even in a much smaller scale, unlike the case of blast furnace because of low capital investment and reduced operating cost.

v) The productivity in terms of $t/m^3/day$ of hot metal is very high as iron ore gets reduced in its solid state as well as in the liquid slag phase.

vi) Increased flexibility in terms of raw materials, scale of operation, operational conditions and product.

vii) The dust generation in the plant gets reduced as sintering and coke oven plants are not necessary.

viii) Many smelting reduction processes employ a fluidized bed for pre-reduction of iron ore because fluidized bed is ideal for the reduction of iron ore[8].

ix) Adequate control of sensible heat and chemical heat in the exit gas which may comply with the energy requirement at the steel works[21].

This process are classified into three types i.e. single stage, two stage and three stage coal-oxygen processes. Among these single stage configuration needs the highest energy. Two stage smelting reduction processes has been found to be most techno-economic viable. Of the various processes based on two stage smelting reduction concept, COREX-process has been found to be commercially suitable and few plants have been installed worldwide[9][3]. Development of this energy efficient, highly productive and economically attractive process

requires understanding of the factor that control degree of post combustion, heat transfer, slag foaming and reduction kinetics amongst other factors. To be energy efficient, it is necessary to have a high degree of post combustion and a relatively low degree of pre-reduction. Combination of low post combustion and high degree of pre-reduction leads to wastage of excess process gas energy, more than that required in the various units in an integrated plant. Heat generated by post combustion transfers itself from hot gas to slag and/or from slag to hot metal. Post combustion in a foamed slag can yield a higher heat transfer efficiency[4].

Since in the current ironmaking scenario the two stage smelting reduction process has gained a lot of importance. The objective of the present work is to model such a process thermochemically so as to get information about it. Main smelting reduction processes are discussed in chapter two. The chapter emphasizes the emerging need for the COREX process which is modeled in chapter three. Chapter four discusses the results obtained from the model. Finally chapter five outlines the conclusions and chapter six scope for the further work in this area.

Chapter 2

Literature Survey

The steel industry, worldwide needed a true alternative for the blast furnace. Because blast furnace requires coke and pelletized iron ore for successful operation. The coking coal reserves are scarce and environmental pollution comes from sintering and pelletization plant and coke oven. Therefore it has been repeatedly tried during past few years to produce a similar product of blast furnace by avoiding these problems.[10]

Few years ago, natural gas based direct reduction technology began to compete with blast furnace. But today these processes are no longer under consideration in the industrialized countries and in areas where high amount of natural gas are still being flared, the demand for sponge iron has remained small. The sponge iron production on the basis of coal was intended to replace natural gas and coke. But these processes are not considered for alternative because sponge iron always comes off badly compared with hot metal as far as its value is concerned. This is not only connected with the lack of perceptible heat and/or melting heat but also with the separation from the gangue. So, it needs proper handling.[10]

Today, smelting reduction has come as true alternative for blast furnace in which liquid iron is obtained by reduction of iron ore at high temperature by using coal and oxygen with a substantial part of the reduction taking place in the liquid oxide phase to increase transport rate due to the enlargement of specific phase contact areas[4][6]. This process employs a fluidized bed for pre-reduction of iron ore so that agglomeration of burden is not required.[8]

2.1 Thermochemical Configurations For Smelting Reduction

The following thermo-chemical configurations can be realized for smelting reduction based on the flow of materials and heat.

1. Single stage process
2. Two stage process
3. Three stage process

2.1.1 Single Stage Process

The simplest reactor configuration is a single well stirred vessel in which cold ore, fuel and oxygen are reacted (see figure 2.1) and both reduction and smelting process takes place. Since the vessel is fed with cold ore without pre-reduction, a higher amount of energy with respect to more efficient process is required, which is met by burning coal with oxygen with a high degree of post combustion[3].

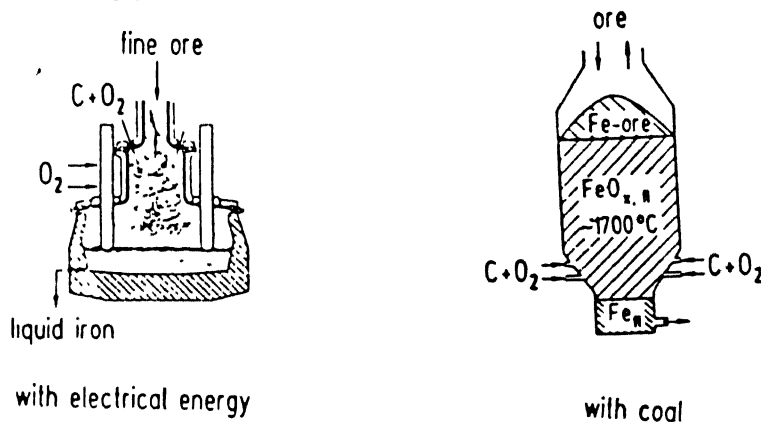


Figure 2.1: Principles of single stage smelting reduction process

The fuel requirement and top gas arising with various fuels are given in table 2.1. This table shows that the single stage smelting reduction process is extremely inefficient and is not technically or, economically attractive for the production of iron unless adequate and large amount of top gas are obtained. Foamy slag and refractory erosion are two other associated problems. The main advantages are low capital cost, use of ore and coal without pretreatment and acceptability of low grade ore and coal.

Table 2.1: Top gas sensible heat and calorific value for various fuels[9]

Fuel (Dry ash free)	Top gas	
	Sensible heat (GJ)	Calorific value (GJ)
Carbon	8.40	43.9
Bituminous coal	12.4	63.6
Subbituminous coal	16.4	84.0

2.1.2 Two Stage Process

The improvement to the single stage process is to utilize the melting offgases for pre-reduction, thereby reducing the energy requirement in the melter to that for melting and final reduction and usefully employing part of the chemical energy in the offgases within the process. This concept has been shown in figure 2.2.

In an integrated process of this nature it is fundamental to match the performance of the two operations. One possible way of operating the process that the offgas make matches the requirement to produce highly metallized ore[11]. For reduction to high metallization level using a reducing gas, the energy content of this reducing gas should be too high because of the thermodynamic and process constraints which apply during the reduction of iron ore. Since the utilization of CO and H_2 for wustite-iron equilibrium at $1200^\circ K$ is 30% and 37% respectively[12]. The top gas will still have very high calorific value[10]. Since pure oxygen is used for coal combustion here, very less amount of N_2 coming from coal will be available

in the top gas. The melter gasifier generates some surplus gas depending on the requirement of reducing

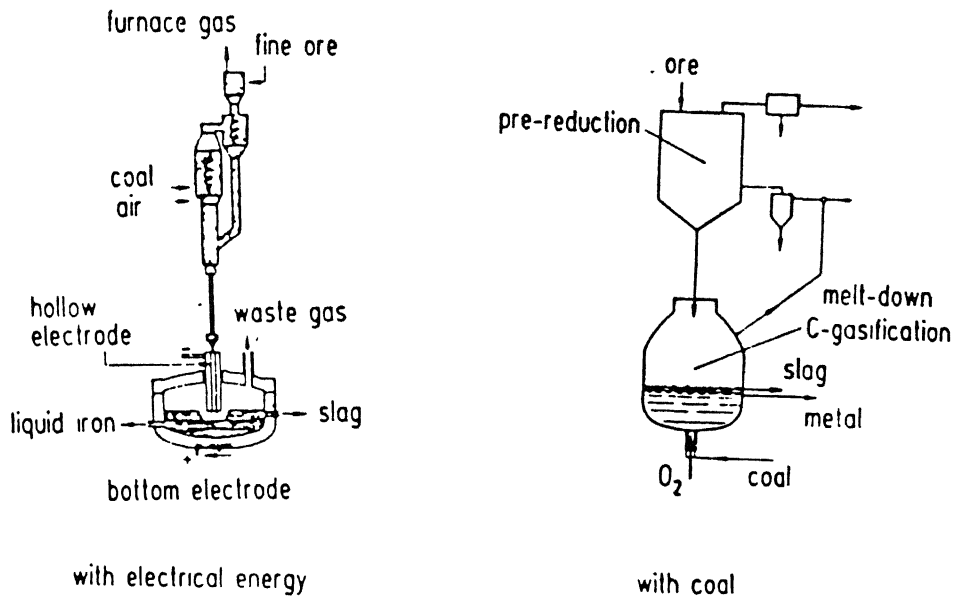


Figure 2.2: Principles of two stage smelting reduction process

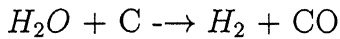
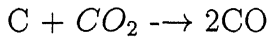
gases in the reduction shaft furnace. This surplus gas mixed with top gas and resulting gas mixture is export gas contains high calorific value. The rejection of energy in the export gas can be avoided by the removal of CO_2 and H_2O from the export gas and recycling of these gases through the reduction unit. By these means it is possible to conserve the chemical energy value of the reducing species and compress and reheat the recycled gas. It is worth noting that the energy consumption for the recycle operation is significantly greater than once through process under conditions where the top gas energy is used effectively. So, it is advantageous to produce DRI in a separate facility by utilizing the top gas for the reduction work[3].

It is clear that energy consumption for the two stage process is significantly superior to the single stage one described above. However, the process contains two inherent sources of heat rejection. The first is the requirement to cool the melter offgases and the second is

that from top gas exited from the reduction shaft furnace. The three stage process aims to utilize these sources of heat[9].

2.1.3 Three Stage Process

The interposition of a reaction zone containing free carbon between the melting and reduction zone provides a means of reducing the temperature of the melter offgas, without rejection of energy through the reactions :



These reactions are both highly endothermic and effectively provide a mechanism whereby excess sensible heat in the gas can be converted to chemical energy available for reduction. Obviously, for these reactions to proceed it is necessary for the gases to contain some oxidized species, although the amount needed are not large. The thermodynamics of the carbon-oxygen system impose a lower limit of temperature which can be reached by the gasification reaction. The various types of three stage processes are shown in figure 2.3.

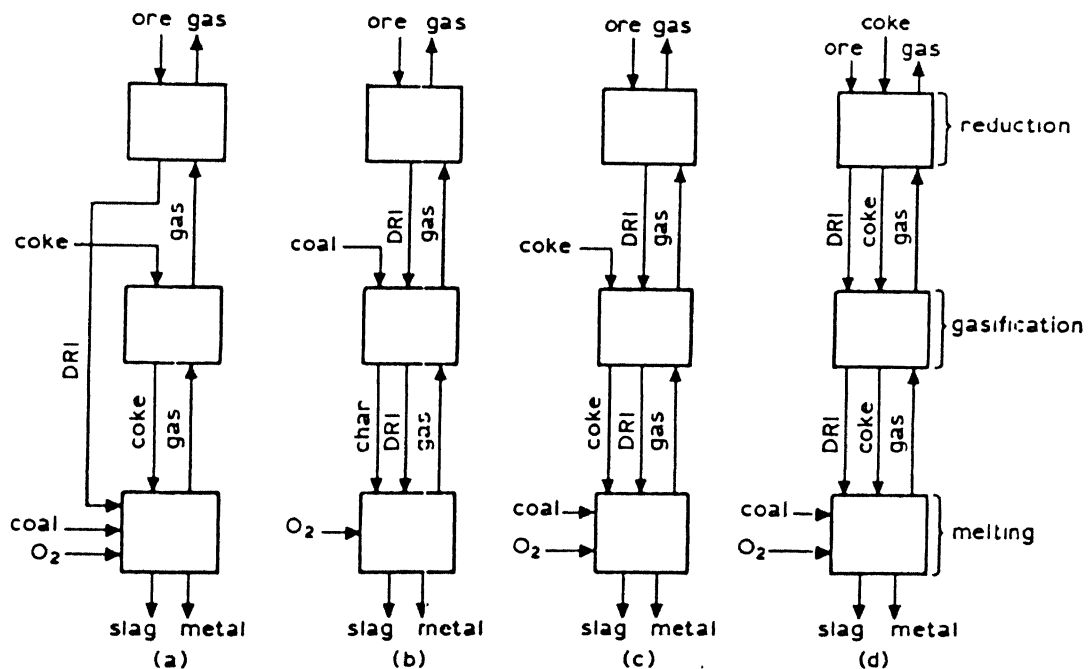


Figure 2.3: Principles of three stage smelting reduction process

The third stage typically consists of a bed of carbonaceous material and in practice it is not necessarily separated from the melter reactor and in melters based on a coke bed the gasification reactions will take part in the upper portion of the melter reactor[9]. The configuration (a) in figure 2.3 is essentially that of Kawasaki and Plasma-smelt process. Configuration (b) is the Korf-KR process, while the Sumitomo-SC process is effectively depicted by configuration (c). In the final configuration (d) the carbon required to make up the consumption in the gasification zone is introduced into the reduction zone. This fully countercurrent operation addresses the second heat rejection area of the two stage process by utilizing the excess sensible heat in the reduction top gases to preheat the incoming coke. This configuration is achieved by the BSC oxy-coal blast furnace.

Thus, the incorporation of third stage results in a process with potentially the lowest energy consumption. However, there are of course energy inefficiencies associated with production of coke which is used in the majority of three stage process.

2.2 Techno-economic Analysis

The energy consumption in iron production for all three configuration has been given in table 2.2. Single stage process is significantly worse than the other process and this is only of interest where there is a local customer with a large requirement for high calorific value gas.

A dramatic improvement is achieved by the two stage process by utilizing the reduction gas within the process itself. Three stage process offers lower energy consumption[8]. But the biggest disadvantage of this process is using coke. Because of scarcity of coking coal, making coke oven and environmental problems, three stage smelting reduction process can not be taken as techno-economic viable in the current ironmaking scenario.

The energy cost for three stage process utilizing coke are somewhat higher than the two stage process because of the cost difference between coal and coke. So, the only process

Table 2.2: Energy consumption for three configurations and the blast furnace for iron production(GJ/THM)

Process type	Net energy consumption
Single stage	26.5 - 29.9
Two stage	15.5 - 16.5
Three stage	
a and c	13.9 - 14.5
b	13.5 - 13.9
d	13.3 - 13.6
Blast furnace	14.7

techno-economic viable is two stage smelting reduction process.

Various smelting reduction processes can be categorized into the three stages described above. This categorization has been shown in table 2.3.

Table 2.3: Smelting reduction processes[9]

Single Stage Process	Two Stage Process	Three stage Process
Coin BSC Hoogovens Hismelt CIG KOBE BSC-ICI	COREX XR INRED ELRED Combismelt	Kawasaki Plasma Smelt KORF-KR Sumitomo-SC

There is only one operational realization of the two stage smelting reduction process which is the COREX process developed by Voest Alpine of Austria. Other smelting reduction technologies are not commercially viable.

2.3 COREX Process Description

This is the first and so far the only commercialized smelting reduction process. Formerly known as the KR process, COREX is the joint development of Korf Engineering and Voest Alpine.

A schematic flowsheet of the process is given in figure 2.4.

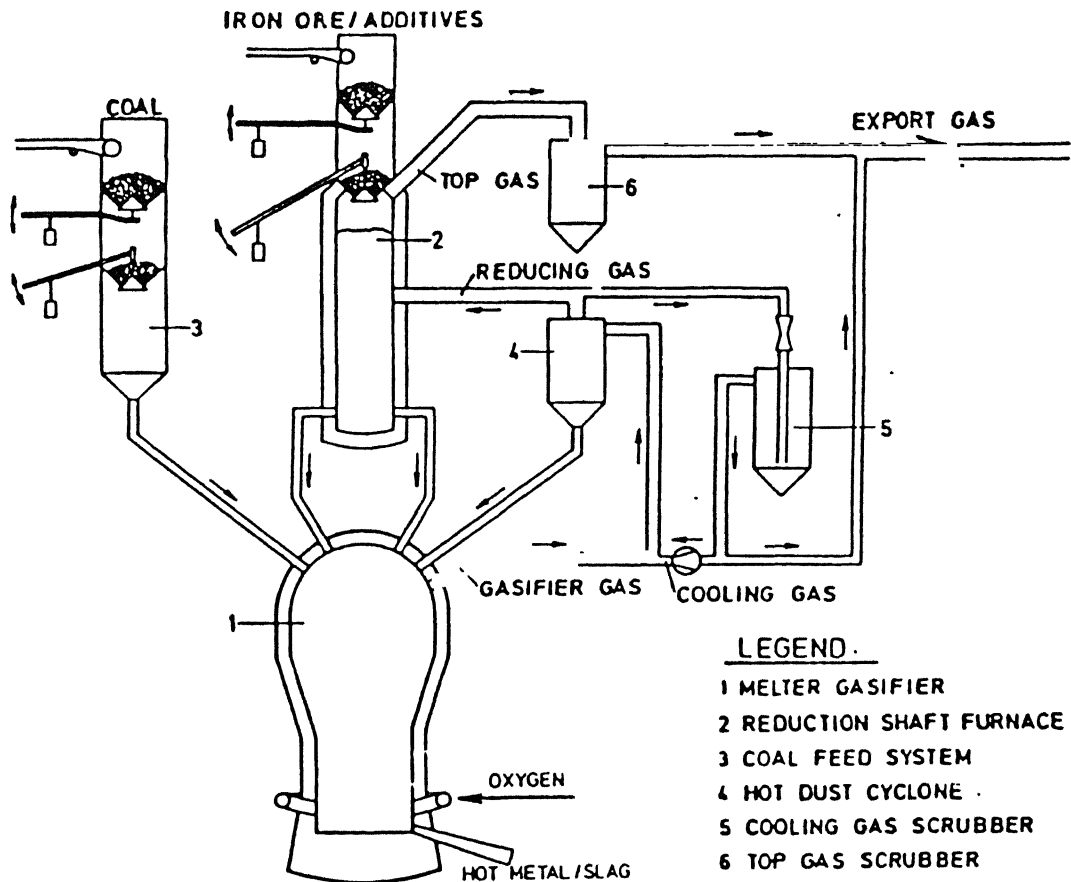


Figure 2.4: COREX Process Flowsheet

It consists of two main units, a fluidized bed melter-gasifier is designed to operate at pressures upto 5 bar[4]. It has a hemispherical top and a cylindrical melting zone and hearth. A bed of coal char is maintained in the melter-gasifier by the addition of raw coal.

Fine ore is first charged into reduction shaft via a lock hopper system where they are

reduced by a reduction gas moving in counter flow. The coal is stored in a pressurized feed bin and charged in the melter-gasifier by a speed controlled feed screw. The coal falls by gravity into the melter-gasifier where it comes into contact with a reducing gas atmosphere at a temperature of approximately 1000°C to 1200°C . Instantaneous drying and degasification of coal particles occurs in the upper portion of the melter gasifier[10].

Generation of reducing gas is done in fluidized bed by oxidation of coal[23]. The heat required in the melter-gasifier for metal and slag melting, residual iron oxide reduction, gas heating and vessel losses is generated in the melt by this oxidation and post combustion[19]. First of all the carbon is oxidized to CO_2 and then CO_2 reacts with fine carbon to form CO. Due to high temperature pre-dominating in the dome of melter-gasifier(in excess of 1000°C), the higher hydrocarbons released from coal during devolatilization and immediately associated to CO and H_2 . Undesirable products such as tars, phenols etc are destroyed and can not be released to atmosphere[7]. After leaving the melter-gasifier, the gas is mixed with cooling gas to attain a temperature about 800°C to 900°C . The gas is then cleaned in hot cyclone[14] and fed to the reduction shaft furnace as reducing gas. The fines captured in the hot cyclone are recirculated into the melter-gasifier via a dust burners[10].

The reducing gas is fed into the reduction shaft furnace through a bustle and ascends through the iron burden according to the counterflow principle[10]. Discharge screw conveys the directly reduced iron from the reduction shaft furnace into the melter-gasifier where final reduction of directly reduced iron takes place. Hot metal and slag drop to the bottom of the melter-gasifier. Analogous to blast furnace hot metal and slag are discharged by conventional tapping procedure.

2.4 COREX vs. BLAST FURNACE

Following comparisons are there between COREX process and blast furnace[3][8][4] :

COREX	BLAST FURNACE
Export gas (high CV)	Top gas (low CV)
Less polluting	More polluting
Low capital cost	High capital cost
Less space requirement	More space requirement
More flexible	Less flexible
No Blast Temperature	Blast Temperature
No injectants through injected O_2	Injectants (Like H_2O , $O_2 - HC$) through air blast
Less energy consumption	High energy consumption
No burden preparation	Coke ovens and sintering
Post combustion	No post combustion

2.5 COREX Plants

There are seven COREX plants[3] in operation feature a total production of nearly 5 million tones of hot metal per year. This is equivalent to approx. 1% of the total output of hot metal.

(a) POSCO Pohang Works (South Korea)

Operating since 1995.

Capacity = 2000 thm/day

(b) ISCOR Pretoria Works (South Africa)

Operating since 1989.

Capacity = 1000 thm/day

(c) JINDAL Toranagallu Works (India)

Two plants have been operating since October 1997 and March 1998 respectively.

Capacity = 1,530,000 thm/annum

(d) Suldanha Steel Works (South Africa)

Operating since 1998.

Capacity = 650,000 thm/annum

(e) HANBO Dangjin Works (Asam Bay, Korea)

Two plants have been operating since 1998.

Total capacity = 2 million thm/annum

2.6 Objective Of The Present Study

In the present study a thermo-chemical model of the COREX process has been prepared. Materials balance and heat balance for melter-gasifier is met by coal combustion. For this a computer program is developed to investigate the influence of degree of metallization, post combustion ratio, heat transfer efficiency, post combustion medium on the following :

- i) Coal rate
- ii) Oxygen required
- iii) Post combustion medium required
- iv) Flux rate
- v) Slag rate
- vi) Export gas amount

For the above calculations, the following input are fixed :

- i) Composition of hot metal, Ore and flux
- ii) Tapping temperature of hot metal and slag
- iii) Reducing and top gas temperature
- iv) Heat losses due to radiation and convection

Chapter 3

THERMO-CHEMICAL MODEL

Thermo-chemical model needs complete heat and materials balance of the process. The model is developed by coupling heat and materials balance equations. First material balance equations are prepared. Then heat demand and supply terms are formulated mathematically and are equated to give heat balance equation.

Objective of the model for two stage smelting reduction process is to provide following information :

- i) Coal consumption
- ii) Ore requirement
- iii) Oxygen requirement for coal combustion
- iv) Medium required for post combustion
- v) Export gas produced
- vi) Flux requirement
- vii) Slag produced

3.1 Model Formulation

3.1.1 Assumptions

Following assumptions are made :

1. Process is operating at steady state.
2. No loss of materials occurs during their transfer from one reactor to the other.
3. Coal combusts completely in the melter-gasifier with the stoichiometric amount of oxygen which is supplied through the side of the melter-gasifier.
4. Iron enters in the system through the oxide of iron i.e. Fe_2O_3 .
5. FeO in the output slag is considered to be negligibly small and neglected.
6. The decomposition of CO in the reduction shaft furnace is considered to be negligibly small and hence neglected.
7. Gases in the melter-gasifier are in thermally and chemically equilibrium with the hot metal containing high percent of carbon. As a consequence of which combustion of coal produces CO and releases hydrogen.
8. Thermal and chemical requirements of the melter-gasifier are completely met by combustion of coal.
9. The rate of conversion of different reactants into products in the melter-gasifier and the reduction shaft furnace are considered to be rapid, once equilibrium conditions are maintained.
10. The oxygen or air or preheated air for post combustion are supplied from the top of the melter-gasifier via a lance. Both complete combustion and mixing of the combustion products with the remaining gas is assumed subsequently. Gas temperature and its composition is assumed to be uniform.
11. The dissociation of CO_2 and H_2O at high temperature makes the process of post combustion incomplete. It is assumed that the dissociation equilibrium is established in the gas phase.

3.1.2 Flow Of The Materials

The flow of materials is shown in Fig 3.1 .

3.2 Materials Balance

Hot metal and slag consists of elements and oxides respectively. Elemental balances for C, Fe, Si, Mn, S, P, H, N, O, Ca, Mg and oxide balance for Al_2O_3 have been carried out. Our basis of calculation is per kg mole of product Fe.

Calculation of n_C^{HM} , n_{Si}^{HM} , n_{Mn}^{HM} , n_S^{HM} , n_P^{HM} and M_I are given in appendix I.

3.2.1 C-Balance

Carbon comes into the process from coal and from $CaCO_3$ and $MgCO_3$ in the flux.

One kg mole of $CaCO_3$ and one kg mole $MgCO_3$ decomposition gives one kg mole of C each.

So,

$$n_C^i = n_C^{coal} + n_{CaCO_3}^L + n_{CaCO_3}^D + n_{MgCO_3}^L + n_{MgCO_3}^D \quad (3.1)$$

Where,

$$n_C^{coal} = \frac{M_C \times p_C^{coal}}{12 \times 100} \quad (3.2)$$

$$n_{CaCO_3}^L = \frac{M_L \times p_{CaCO_3}^L}{100 \times 100} \quad (3.3)$$

$$n_{CaCO_3}^D = \frac{M_D \times p_{CaCO_3}^D}{100 \times 100} \quad (3.4)$$

$$n_{MgCO_3}^L = \frac{M_L \times p_{MgCO_3}^L}{84 \times 100} \quad (3.5)$$

$$n_{MgCO_3}^D = \frac{M_D \times p_{MgCO_3}^D}{84 \times 100} \quad (3.6)$$

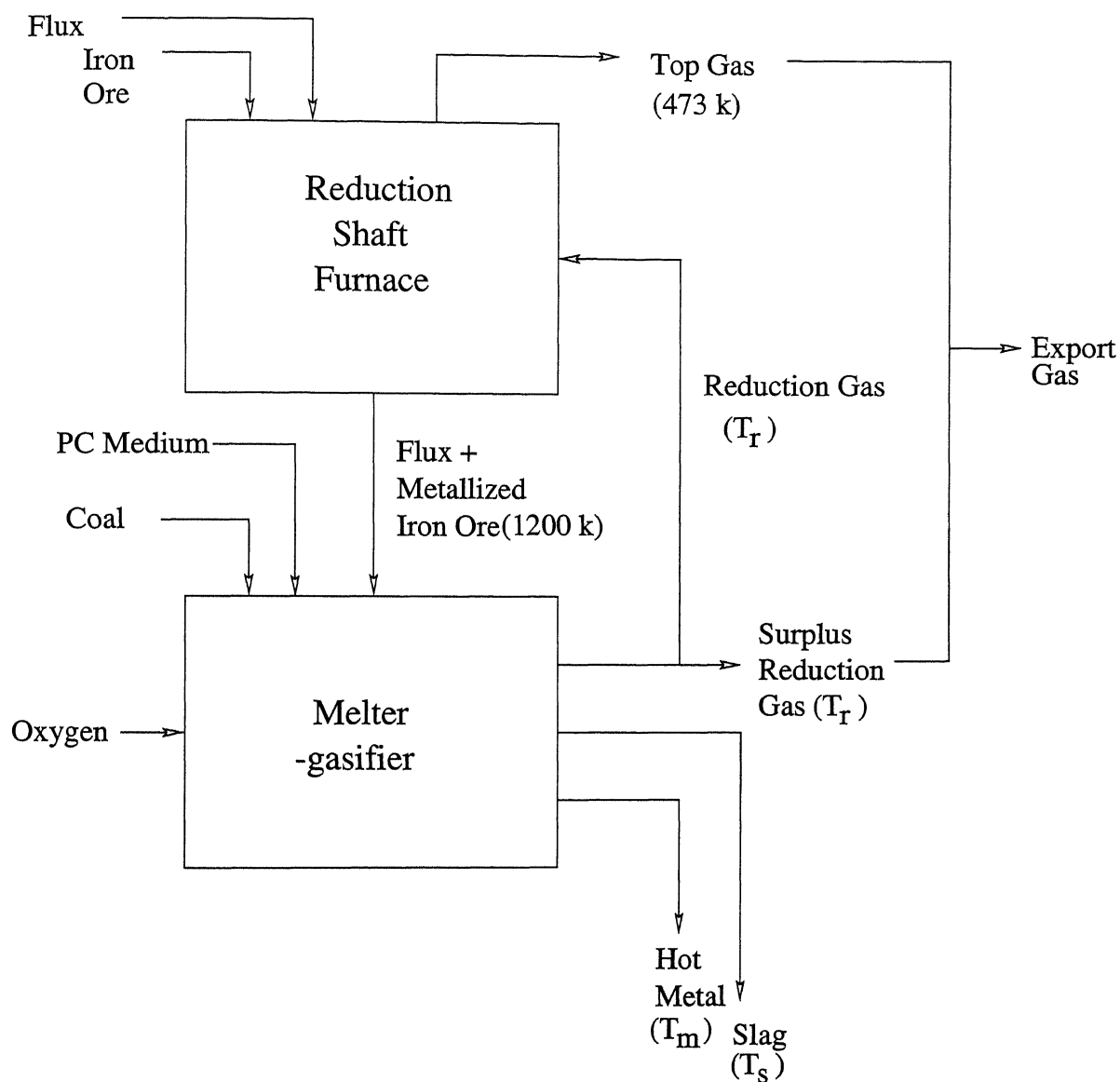


Figure 3.1: Flow of materials in two stage smelting reduction process

So, combining equations (3.1) -(3.6), we get,

$$n_C^i = \frac{M_C \times p_C^{coal}}{12 \times 100} + \frac{M_L \times p_{CaCO_3}^L}{100 \times 100} + \frac{M_D \times p_{CaCO_3}^D}{100 \times 100} + \frac{M_L \times p_{MgCO_3}^L}{84 \times 100} + \frac{M_D \times p_{MgCO_3}^D}{84 \times 100} \quad (3.7)$$

Carbon leaves the process in the hot metal, surplus gas and top gas.

So,

$$n_C^o = n_C^{tg} + n_C^{sg} + n_C^{HM} \quad (3.8)$$

Where,

$$n_C^{tg} = n_{CO}^{tg} + n_{CO_2}^{tg} \quad (3.9)$$

$$n_C^{sg} = n_{CO}^{sg} \times \left(1 - \frac{ppc}{100}\right) + n_{CO_2}^{sg} \times \left(\frac{ppc}{100}\right) \quad (3.10)$$

Where,

$$ppc = \frac{[\%CO_2 + \%H_2O]}{[\%CO + \%CO_2 + \%H_2 + \%H_2O]} \times 100 \quad (3.11)$$

So, combining equations (3.8) - (3.10), we get,

$$n_C^o = n_{CO}^{tg} + n_{CO_2}^{tg} + n_{CO}^{sg} \times \left(1 - \frac{ppc}{100}\right) + n_{CO_2}^{sg} \times \left(\frac{ppc}{100}\right) + n_C^{HM} \quad (3.12)$$

At steady state,

$$n_C^i = n_C^o \quad (3.13)$$

So, from equation (3.7), (3.12) and (3.13), we get

$$\begin{aligned} n_{CO}^{tg} + n_{CO_2}^{tg} + n_{CO}^{sg} \times \left(1 - \frac{ppc}{100}\right) + n_{CO_2}^{sg} \times \left(\frac{ppc}{100}\right) &= \frac{M_C \times p_C^{coal}}{12 \times 100} + \frac{M_L \times p_{CaCO_3}^L}{100 \times 100} \\ &+ \frac{M_D \times p_{CaCO_3}^D}{100 \times 100} + \frac{M_L \times p_{MgCO_3}^L}{84 \times 100} + \frac{M_D \times p_{MgCO_3}^D}{84 \times 100} - n_C^{HM} \end{aligned} \quad (3.14)$$

Active carbon react with oxygen sent for coal combustion, defined as

$$n_C^A = n_{CO}^{tg} + n_{CO_2}^{tg} + n_{CO}^{sg} \times \left(1 - \frac{ppc}{100}\right) + n_{CO_2}^{sg} \times \left(\frac{ppc}{100}\right) \quad (3.15)$$

3.2.2 Fe-Balance

Fe enters the reduction shaft furnace as Fe_2O_3 in the iron ore and leaves as Fe in the hot metal. Since our basis is one kg mole of product Fe,

So,

$$n_{Fe}^i = n_{Fe}^o = 1 \quad (3.16)$$

Iron ore undergoes metallization in the reduction shaft furnace by reduction gas which comes from melter-gasifier. Iron comes out of reduction shaft furnace as $Fe_{0.95}O$ and Fe. $Fe_{0.95}O$ and Fe further changes to liquid iron in melter-gasifier.

i.e.

$$n_{Fe} + 0.95n_{Fe_{0.95}O} = 1 \quad (3.17)$$

Now,

$$X = \frac{n_{Fe} \times 100}{n_{Fe} + 0.95n_{Fe_{0.95}O}} \quad (3.18)$$

Substituting (3.17) in (3.18), we get,

$$n_{Fe} = \left[\frac{X}{100} \right] \quad (3.19)$$

And,

$$n_{Fe_{0.95}O} = \left[\frac{100 - X}{95} \right] \quad (3.20)$$

3.2.3 Si-Balance

Si enters the process as SiO_2 from iron ore, coal ash, limestone and dolomite.

One kg mole of SiO_2 corresponds to one kg mole of Si

So,

$$n_{Si}^i = n_{SiO_2}^{ash} + n_{SiO_2}^{ore} + n_{SiO_2}^L + n_{SiO_2}^D \quad (3.21)$$

Where,

$$n_{SiO_2}^{ash} = \frac{M_C \times p_{ash}^{coal} \times p_{SiO_2}^{ash}}{60 \times 100 \times 100} \quad (3.22)$$

$$n_{SiO_2}^{ore} = \frac{M_I \times p_{SiO_2}^{ore}}{60 \times 100} \quad (3.23)$$

$$n_{SiO_2}^L = \frac{M_L \times p_{SiO_2}^L}{60 \times 100} \quad (3.24)$$

$$n_{SiO_2}^D = \frac{M_D \times p_{SiO_2}^D}{60 \times 100} \quad (3.25)$$

Substituting equations (3.22), (3.23), (3.24), (3.25) in equation (3.21), we get

$$n_{Si}^i = \frac{M_C \times p_{ash}^{coal} \times p_{SiO_2}^{ash}}{60 \times 100 \times 100} + \frac{M_I \times p_{SiO_2}^{ore}}{60 \times 100} + \frac{M_L \times p_{SiO_2}^L}{60 \times 100} + \frac{M_D \times p_{SiO_2}^D}{60 \times 100} \quad (3.26)$$

Silicon leaves the process as Si in the hot metal and as SiO_2 in the slag.

So,

$$n_{Si}^o = n_{Si}^{HM} + n_{SiO_2}^{slag} \quad (3.27)$$

At steady state,

$$n_{Si}^i = n_{Si}^o \quad (3.28)$$

Substituting (3.26), (3.27) in equation (3.28),

$$n_{SiO_2}^{slag} = \frac{M_C \times p_{ash}^{coal} \times p_{SiO_2}^{ash}}{60 \times 100 \times 100} + \frac{M_I \times p_{SiO_2}^{ore}}{60 \times 100} + \frac{M_L \times p_{SiO_2}^L}{60 \times 100} + \frac{M_D \times p_{SiO_2}^D}{60 \times 100} - n_{Si}^{HM} \quad (3.29)$$

3.2.4 Mn-Balance

Mn enters the process as MnO from iron ore. One kg mole of MnO corresponds to one kg mole of Mn.

So,

$$n_{Mn}^i = n_{MnO}^{ore} \quad (3.30)$$

Where,

$$n_{MnO}^{ore} = \frac{M_I \times p_{MnO}^{ore}}{70.9 \times 100} \quad (3.31)$$

Substituting equation (3.31) in equation (3.30),

$$n_{Mn}^i = \frac{M_I \times p_{MnO}^{ore}}{70.9 \times 100} \quad (3.32)$$

Manganese leaves the process as Mn in the hot metal and as MnO in the slag.

So,

$$n_{Mn}^o = n_{Mn}^{HM} + n_{MnO}^{slag} \quad (3.33)$$

At steady state,

$$n_{Mn}^i = n_{Mn}^o \quad (3.34)$$

Substituting (3.32), (3.33) in equation (3.34),

$$n_{MnO}^{slag} = \frac{M_I \times p_{MnO}^{ore}}{70.9 \times 100} - n_{Mn}^{HM} \quad (3.35)$$

3.2.5 S-Balance

Sulphur comes into the process from iron ore as FeS and coal as S.

One kg mole of FeS corresponds to one kg mole of sulphur

So,

$$n_S^i = n_{FeS}^{ore} + n_S^{coal} \quad (3.36)$$

Where,

$$n_{FeS}^{ore} = \frac{M_I \times p_{FeS}^{ore}}{88 \times 100} \quad (3.37)$$

$$n_S^{coal} = \frac{M_C \times p_S^{coal}}{32 \times 100} \quad (3.38)$$

Substituting (3.37) and (3.38) into equation (3.36),

$$n_S^i = \frac{M_I \times p_{FeS}^{ore}}{88 \times 100} + \frac{M_C \times p_S^{coal}}{32 \times 100} \quad (3.39)$$

Sulphur leaves the process in the hot metal as S and in the slag as CaS.

One kg mole of CaS corresponds to one kg mole of S.

So,

$$n_S^o = n_S^{HM} + n_{CaS}^{slag} \quad (3.40)$$

At steady state,

$$n_S^i = n_S^o \quad (3.41)$$

Substituting equations (3.39) and (3.40) into equation (3.41),

$$n_{CaS}^{slag} = \frac{M_I \times p_{FeS}^{ore}}{88 \times 100} + \frac{M_C \times p_S^{coal}}{32 \times 100} - n_S^{HM} \quad (3.42)$$

3.2.6 P-Balance

Phosphorus enters the process from iron ore as P_2O_5 . One kg mole of P_2O_5 corresponds to two kg moles of phosphorus.

So,

$$n_P^i = 2n_{P_2O_5}^{ore} \quad (3.43)$$

Where,

$$n_{P_2O_5}^{ore} = \frac{M_I \times p_{P_2O_5}^{ore}}{142 \times 100} \quad (3.44)$$

Substituting equation (3.44) into equation (3.43),

$$n_P^i = \frac{2 \times M_I \times p_{P_2O_5}^{ore}}{142 \times 100} \quad (3.45)$$

Phosphorus leaves the process as P in the hot metal.

So,

$$n_P^o = n_P^{HM} \quad (3.46)$$

At steady state,

$$n_P^i = n_P^o \quad (3.47)$$

Substituting equations (3.45) and (3.46) into equation (3.47),

$$\frac{2 \times M_I \times p_{P_2O_5}^{ore}}{142 \times 100} = n_P^{HM} \quad (3.48)$$

3.2.7 H-Balance

Hydrogen enters the process from volatile matter and moisture in coal.

So,

$$n_H^i = n_H^{coal} + n_H^M \quad (3.49)$$

Where,

$$n_H^{coal} = \frac{M_C \times p_H^{coal}}{100} \quad (3.50)$$

$$n_H^M = \frac{2 \times M_C \times p_M^{coal}}{18 \times 100} \quad (3.51)$$

Substituting equations (3.50) and (3.51) in equation (3.49),

$$n_H^i = \frac{M_C \times p_H^{coal}}{100} + \frac{2 \times M_C \times p_M^{coal}}{18 \times 100} \quad (3.52)$$

Hydrogen leaves the process as H_2 and H_2O in the surplus and top gas. One kg mole of H_2 and H_2O both corresponds to two kg moles of H.

So,

$$n_H^o = n_H^{sg} + n_H^{tg} \quad (3.53)$$

Where,

$$n_H^{tg} = 2 \times [n_{H_2}^{tg} + n_{H_2O}^{tg}] \quad (3.54)$$

and,

$$n_H^{sg} = 2 \times \left[n_{H_2}^{sg} \times \left(1 - \frac{ppc}{100} \right) + n_{H_2O}^{sg} \times \frac{ppc}{100} \right] \quad (3.55)$$

Substituting equations (3.54) and (3.55) into equation (3.53),

$$n_H^o = 2 \times [n_{H_2}^{tg} + n_{H_2O}^{tg}] + 2 \times \left[n_{H_2}^{sg} \times \left(1 - \frac{ppc}{100} \right) + n_{H_2O}^{sg} \times \frac{ppc}{100} \right] \quad (3.56)$$

At steady state,

$$n_H^i = n_H^o \quad (3.57)$$

Substituting equations (3.52) and (3.56) into equation (3.57),

$$2 \times [n_{H_2}^{tg} + n_{H_2O}^{tg}] + 2 \times \left[n_{H_2}^{sg} \times \left(1 - \frac{ppc}{100} \right) + n_{H_2O}^{sg} \times \frac{ppc}{100} \right] = \frac{M_C \times p_H^{coal}}{100} + \frac{2 \times M_C \times p_M^{coal}}{18 \times 100} \quad (3.58)$$

3.2.8 Ca-Balance

Calcium enters the process from limestone and dolomite as $CaCO_3$ and from ash as CaO. one kg mole of $CaCO_3$ and CaO corresponds to one kg mole of Ca each.

So,

$$n_{Ca}^i = n_{CaO}^{ash} + n_{CaCO_3}^L + n_{CaCO_3}^D \quad (3.59)$$

Where,

$$n_{CaO}^{ash} = \frac{M_C \times p_{ash}^{coal} \times p_{CaO}^{ash}}{56 \times 100 \times 100} \quad (3.60)$$

Substituting equation (3.60) into equation (3.59),

$$n_{Ca}^i = \frac{M_C \times p_{ash}^{coal} \times p_{CaO}^{ash}}{56 \times 100 \times 100} + n_{CaCO_3}^L + n_{CaCO_3}^D \quad (3.61)$$

Calcium leaves the process in the slag as CaO and CaS. One kg mole of CaS corresponds to one kg mole of Ca.

So,

$$n_{Ca}^o = n_{CaO}^{slag} + n_{CaS}^{slag} \quad (3.62)$$

At steady state,

$$n_{Ca}^i = n_{Ca}^o \quad (3.63)$$

Substituting equations (3.61) and (3.62) into equation (3.63),

$$n_{CaO}^{slag} = \frac{M_C \times p_{ash}^{coal} \times p_{CaO}^{ash}}{56 \times 100 \times 100} + n_{CaCO_3}^L + n_{CaCO_3}^D - n_{CaS}^{slag} \quad (3.64)$$

Where, $n_{CaCO_3}^L$, $n_{CaCO_3}^D$ and n_{CaS}^{slag} are taken from equations (3.3), (3.4) and (3.42) respectively.

3.2.9 O-Balance

Oxygen enters the process as oxygen for coal combustion, for post combustion, from ore as Fe_2O_3 , SiO_2 , MnO , P_2O_5 , from limestone as $CaCO_3$ and $MgCO_3$, from coal moisture and coal oxygen.

So,

$$\begin{aligned} n_O^i &= n_O^B + n_O^{pc} + [O/Fe] + 2 \times [Si/Fe] + [Mn/Fe] + 2.5 \times [P/Fe] \\ &+ 2 \left[n_{CaCO_3}^L + n_{CaCO_3}^D + n_{MgCO_3}^L + n_{MgCO_3}^D \right] + n_O^M + n_O^{coal} \end{aligned} \quad (3.65)$$

Where,

$$n_O^M = \frac{M_C \times p_M^{coal}}{18 \times 100} \quad (3.66)$$

$$n_O^{coal} = \frac{M_C \times p_O^{coal}}{16 \times 100} \quad (3.67)$$

Since, iron oxide enters the process as Fe_2O_3 ,

So, $[O/Fe] = 1.5$

Substituting equations (3.66) and (3.67) into equation (3.65),

$$\begin{aligned} n_O^i &= n_O^B + n_O^{pc} + 1.5 + 2 \times [Si/Fe] + [Mn/Fe] + 2.5 \times [P/Fe] \\ &+ 2 \left[n_{CaCO_3}^L + n_{CaCO_3}^D + n_{MgCO_3}^L + n_{MgCO_3}^D \right] \\ &+ \frac{M_C \times p_M^{coal}}{18 \times 100} + \frac{M_C \times p_O^{coal}}{16 \times 100} \end{aligned} \quad (3.68)$$

Oxygen leaves the process in the surplus gas and the top gas as CO , CO_2 and H_2O .

So,

$$n_O^o = [O/C]^{sg}.n_C^A + [O/H_2]^{sg}.n_{H_2}^o + [O/C]^{tg}.n_C^A + [O/H_2]^{tg}.n_{H_2}^o \quad (3.69)$$

Where,

$$[O/C]^{sg} = \frac{n_{CO}^{sg} + 2n_{CO_2}^{sg}}{n_{CO}^{sg} + n_{CO_2}^{sg}} \quad (3.70)$$

$$[O/C]^{tg} = \frac{n_{CO}^{tg} + 2n_{CO_2}^{tg}}{n_{CO}^{tg} + n_{CO_2}^{tg}} \quad (3.71)$$

$$[O/H_2]^{sg} = \frac{n_{H_2O}^{sg}}{n_{H_2}^{sg} + n_{H_2O}^{sg}} \quad (3.72)$$

$$[O/H_2]^{tg} = \frac{n_{H_2O}^{tg}}{n_{H_2}^{tg} + n_{H_2O}^{tg}} \quad (3.73)$$

and,

$$n_H^o = 2n_{H_2}^o \quad (3.74)$$

Substituting equations (3.70) - (3.74) into equation (3.69),

$$n_O^o = \frac{n_{CO}^{sg} + 2n_{CO_2}^{sg}}{n_{CO}^{sg} + n_{CO_2}^{sg}}.n_C^A + \frac{n_{H_2O}^{sg}}{n_{H_2}^{sg} + n_{H_2O}^{sg}}.\frac{1}{2}n_H^o + \frac{n_{CO}^{tg} + 2n_{CO_2}^{tg}}{n_{CO}^{tg} + n_{CO_2}^{tg}}.n_C^A + \frac{n_{H_2O}^{tg}}{n_{H_2}^{tg} + n_{H_2O}^{tg}}.\frac{1}{2}n_H^o \quad (3.75)$$

n_C^A and n_H^o are taken from equations (3.15) and (3.56) respectively.

Since post combustion of one kg mole of CO to CO_2 and one kg mole of H_2 to H_2O both requires one kg mole of O each.

So,

$$n_O^{pc} = n_{CO_2}^r + n_{H_2O}^r \quad (3.76)$$

Where,

$$n_{CO_2}^r = n_C^A \cdot \frac{ppc}{100} \quad (3.77)$$

$$n_{H_2O}^r = \frac{1}{2} n_H^o \cdot \frac{ppc}{100} \quad (3.78)$$

Substituting equations (3.77) and (3.78) into equation (3.76),

$$n_O^{pc} = n_C^A \cdot \frac{ppc}{100} + \frac{1}{2} n_H^o \cdot \frac{ppc}{100} \quad (3.79)$$

At steady state,

$$n_O^i = n_O^o \quad (3.80)$$

Combining equations (3.68), (3.75), (3.79) and (3.80), we get

$$\begin{aligned} n_O^B = & \frac{n_{CO}^{sg} + 2n_{CO_2}^{sg}}{n_{CO}^{sg} + n_{CO_2}^{sg}} \cdot n_C^A + \frac{n_{CO}^{tg} + 2n_{CO_2}^{tg}}{n_{CO}^{tg} + n_{CO_2}^{tg}} \cdot n_C^A + \frac{n_{H_2O}^{sg}}{n_{H_2}^{sg} + n_{H_2O}^{sg}} \cdot \frac{1}{2} n_H^o + \frac{n_{H_2O}^{tg}}{n_{H_2}^{tg} + n_{H_2O}^{tg}} \cdot \frac{1}{2} n_H^o \\ & - n_C^A \cdot \frac{ppc}{100} - \frac{1}{2} n_H^o \cdot \frac{ppc}{100} - 1.5 - 2 \left[n_{CaCO_3}^L + n_{CaCO_3}^D + n_{MgCO_3}^L + n_{MgCO_3}^D \right] \\ & - 2[Si/Fe] - [Mn/Fe] - 2.5[P/Fe] - \frac{M_C \times p_M^{coal}}{18 \times 100} - \frac{M_C \times p_O^{coal}}{16 \times 100} \end{aligned} \quad (3.81)$$

3.2.10 N-Balance

Nitrogen comes into the process from coal and post combustion medium like air or, O_2 -enriched air.

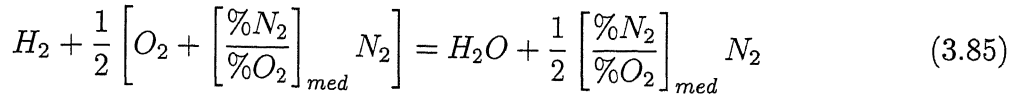
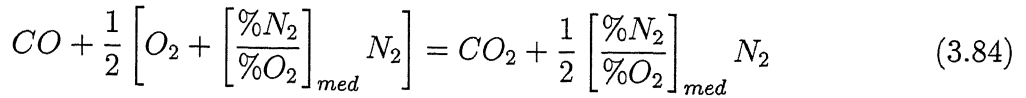
So,

$$n_N^i = n_N^{coal} + n_N^{pc} \quad (3.82)$$

Where,

$$n_N^{coal} = \frac{M_C \times p_N^{coal}}{14 \times 100} \quad (3.83)$$

For post combustion, these two reactions occur :



Since post combustion of one kg mole of CO to CO_2 and one kg mole H_2 to H_2O both gives $\left[\frac{\%N_2}{\%O_2} \right]_{med}$ kg mole of N each.

So,

$$n_N^{pc} = \left[\frac{\%N_2}{\%O_2} \right]_{med} \cdot [n_{CO_2}^r + n_{H_2O}^r] \quad (3.86)$$

Substituting equations (3.83) and (3.86) into equation (3.82),

$$n_N^i = \frac{M_C \times p_N^{coal}}{14 \times 100} + \left[\frac{\%N_2}{\%O_2} \right]_{med} \cdot [n_{CO_2}^r + n_{H_2O}^r] \quad (3.87)$$

When oxygen is used for post combustion, then $\left[\frac{\%N_2}{\%O_2} \right]_{med}$ is zero.

Nitrogen leaves the process as N_2 in the surplus and top gas.

So,

$$n_N^o = n_N^{sg} + n_N^{tg} \quad (3.88)$$

At steady state,

$$n_N^i = n_N^o \quad (3.89)$$

So, by combining equations (3.87) - (3.89), we get,

$$n_N^{sg} + n_N^{tg} = \frac{M_C \times p_N^{coal}}{2 \times 14 \times 100} \quad (3.90)$$

3.2.11 Mg-Balance

Magnesium enters the process from limestone and dolomite as $MgCO_3$ and from ash as MgO . One kg mole of $MgCO_3$ gives one kg mole of Mg and one kg mole of MgO gives one kg mole of Mg.

So,

$$n_{Mg}^i = n_{MgO}^{ash} + n_{MgCO_3}^L + n_{MgCO_3}^D \quad (3.91)$$

Where,

$$n_{MgO}^{ash} = \frac{M_C \times p_{ash}^{coal} \times p_{MgO}^{ash}}{40 \times 100 \times 100} \quad (3.92)$$

Magnesium leaves the process in the slag as MgO .

So,

$$n_{Mg}^o = n_{MgO}^{slag} \quad (3.93)$$

At steady state,

$$n_{Mg}^i = n_{Mg}^o \quad (3.94)$$

So, combining equations (3.91)- (3.94), we get,

$$n_{MgO}^{slag} = \frac{M_C \times p_{ash}^{coal} \times p_{MgO}^{ash}}{40 \times 100 \times 100} + n_{MgCO_3}^L + n_{MgCO_3}^D \quad (3.95)$$

3.2.12 Al_2O_3 -Balance

Al_2O_3 enters the process from iron ore, ash of coal and limestone and dolomite.

So,

$$n_{Al_2O_3}^i = n_{Al_2O_3}^{ore} + n_{Al_2O_3}^{ash} + n_{Al_2O_3}^L + n_{Al_2O_3}^D \quad (3.96)$$

Where,

$$n_{Al_2O_3}^{ore} = \frac{M_I \times p_{Al_2O_3}^{ore}}{102 \times 100} \quad (3.97)$$

$$n_{Al_2O_3}^{ash} = \frac{M_C \times p_{ash}^{coal} \times p_{Al_2O_3}^{ash}}{102 \times 100 \times 100} \quad (3.98)$$

$$n_{Al_2O_3}^L = \frac{M_L \times p_{Al_2O_3}^L}{102 \times 100} \quad (3.99)$$

$$n_{Al_2O_3}^D = \frac{M_D \times p_{Al_2O_3}^D}{102 \times 100} \quad (3.100)$$

Al_2O_3 leaves the process as Al_2O_3 in the slag.

So,

$$n_{Al_2O_3}^o = n_{Al_2O_3}^{slag} \quad (3.101)$$

At steady state,

$$n_{Al_2O_3}^i = n_{Al_2O_3}^o \quad (3.102)$$

From equation (3.96), (3.101) and (3.102), we get,

$$n_{Al_2O_3}^{slag} = n_{Al_2O_3}^{ore} + n_{Al_2O_3}^{ash} + n_{Al_2O_3}^L + n_{Al_2O_3}^D \quad (3.103)$$

Substituting equations (3.97) - (3.100) into equation (3.102),

$$n_{Al_2O_3}^{slag} = \frac{M_I \times p_{Al_2O_3}^{ore}}{102 \times 100} + \frac{M_C \times p_{ash}^{coal} \times p_{Al_2O_3}^{ash}}{102 \times 100 \times 100} + \frac{M_L \times p_{Al_2O_3}^L}{102 \times 100} + \frac{M_D \times p_{Al_2O_3}^D}{102 \times 100} \quad (3.104)$$

3.3 Heat Balance

For steady state, heat accumulation is zero.

So, the equation of heat balance is,

$$\text{Heat demand} = \text{Heat supply} \quad (3.105)$$

3.3.1 Heat Demand

Here heat demand for melter-gasifier and reduction shaft furnace have been calculated. All terms are expressed in KJ/Kg moles of product Fe. The values for enthalpy of formation and equations used to calculate enthalpy as function of temperature are given in appendix III and IV respectively.

3.3.1.1 Melter-Gasifier

The following metallurgical operations are considered for calculating the heat demand for the melter-gasifier.

$$\begin{aligned} [\text{Heat demand}] &= [\text{Hot metal heat demand}] \\ &+ [\text{Slag heat demand}] \\ &+ [\text{Heat absorbed by endothermic reactions}] \\ &+ [\text{Drying of coal}] \\ &+ [\text{Sensible heat of the gas leaving melter-gasifier}] \\ &+ [\text{Latent heat of devolatilization of coal}] \\ &+ [\text{Heat losses}] \end{aligned} \quad (3.106)$$

(a) Hot Metal Heat Demand

Hot metal and slag are considered to be tapped at $T_M^\circ\text{K}$ where T_M is the temperature above the melting point of hot metal and slag.

The amount of heat in hot metal,

$$\Delta H_{HM} = \Delta H_{HM}^S + \Delta H_{HM}^M \quad (3.107)$$

Where, ΔH_{HM}^S is the sensible heat of constituents of hot metal and ΔH_{HM}^M is the heat of mixing of C, Si, Mn, S, P in hot metal.

(i) Sensible heat is the heat required to raise the temperature of various elements Fe, Si, Mn, S and P from $1200^\circ K$ to $T_M^\circ K$ comes from ore and flux through reduction shaft furnace and rest Si, S, C from $298^\circ K$ to $T_M^\circ K$ comes from coal.

Sensible heat for iron which is coming through reduction shaft furnace at $1200^\circ K$ and comes out at $T_M^\circ K$

$$\begin{aligned} &= n_{Fe(HM)} [H_{T_M}^\circ - H_{1200}^\circ] \\ &= n_{Fe(HM)} [(H_{T_M}^\circ - H_{298}^\circ) - (H_{1200}^\circ - H_{298}^\circ)] \\ &= n_{Fe(HM)} [(44T_M - 5800) - (38.4 \times 1200 - 11400)] \\ &= n_{Fe(HM)} [44T_M - 40480] \end{aligned}$$

So, sensible heat,

$$\begin{aligned} \Delta H_{HM}^S &= n_{Fe(HM)} [44T_M - 40480] + [C/Fe] [23.5T_M - 11800] \\ &\quad + [n_{Si}^a + n_{Si}^f] [27.2T_M + 17960] + n_{Si}^c [27.2T_M + 40400] \\ &\quad + [Mn/Fe] [48.6T_M - 42640] + n_S^a [18.7T_M - 22440] \\ &\quad + n_S^c [18.7T_M - 6250] + [P/Fe] [18.65T_M - 22380] \end{aligned} \quad (3.108)$$

As our basis of calculation is per kg mole of product Fe. so, the value of $n_{Fe(HM)}$ is 1.

The calculation of n_{Si}^a , n_{Si}^c , n_{Si}^f , n_S^a and n_S^c are given in appendix V.

(ii) Heat of mixing,

$$\begin{aligned} \Delta H_{HM}^M &= [C/Fe] [30000] + [Si/Fe] [-95000] + [Mn/Fe] [4000] \\ &\quad + [S/Fe] [-132384] + [P/Fe] [-122640] \end{aligned} \quad (3.109)$$

So, combining equations (3.107) - (3.109), we get,

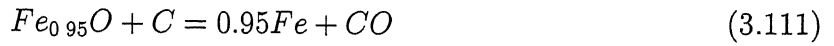
Hot metal heat demand,

$$\Delta H_{HM}^M = [44T_M - 40480] + [C/Fe] [23.5T_M - 11800] + [n_{Si}^a + n_{Si}^f] [27.2T_M + 17960]$$

$$\begin{aligned}
& +n_{S_i}^e [27.2T_M + 40400] + [Mn/Fe] [48.6T_M - 42640] + n_S^a [18.7T_M - 22440] \\
& +n_S^e [18.7T_M - 6250] + [P/Fe] [18.65T_M - 22380] + [C/Fe] [30000] \\
& + [Si/Fe] [-95000] + [Mn/Fe] [4000] + [S/Fe] [-132384] \\
& + [P/Fe] [-122640]
\end{aligned} \tag{3.110}$$

(b) Heat Absorbed By Endothermic Reactions

following endothermic reactions occurs in the melter-gasifier absorbs heat :



Heat required for reaction (3.111),

$$\Delta H_{3.111} = n_{Fe_{0.95}O} [265000] \tag{3.117}$$

Heat required for reaction (3.112),

$$\Delta H_{3.112} = [Mn/Fe] [386000] \tag{3.118}$$

Heat required for reaction (3.113),

$$\Delta H_{3.113} = [n_{S_i}^a + n_{S_i}^f] [901000] + n_{S_i}^c [911000] \tag{3.119}$$

Heat required for reaction (3.114),

$$\Delta H_{3\ 114} = \frac{1}{2} [P/Fe] [1554000] \quad (3.120)$$

Heat required for reaction (3.115),

$$\Delta H_{3\ 115} = n_{CaCO_3}^{flux} [561000] \quad (3.121)$$

Where,

$$n_{CaCO_3}^{flux} = n_{CaCO_3}^L + n_{CaCO_3}^D \quad (3.122)$$

and heat required for reaction (3.116),

$$\Delta H_{3\ 116} = n_{MgCO_3}^{flux} [511000] \quad (3.123)$$

Where,

$$n_{MgCO_3}^{flux} = n_{MgCO_3}^L + n_{MgCO_3}^D \quad (3.124)$$

Heat absorbed by endothermic reactions is the addition of equations (3.117)-(3.121) and (3.124),

$$\begin{aligned} \Delta H_E &= \Delta H_{3.111} + \Delta H_{3\ 112} + \Delta H_{3\ 113} + \Delta H_{3.114} + \Delta H_{3.115} + \Delta H_{3.116} \\ &= n_{Fe_{0.95}O} [265000] + [Mn/Fe] [386000] + [n_{Si}^a + n_{Si}^f] [901000] \\ &\quad + \frac{1}{2} [P/Fe] [1554000] + [n_{CaCO_3}^L + n_{CaCO_3}^D] [561000] \\ &\quad + n_{Si}^c [911000] + [n_{MgCO_3}^L + n_{MgCO_3}^D] [511000] \end{aligned} \quad (3.125)$$

(c) Slag Heat Demand

Slag contains CaO, SiO_2 , Al_2O_3 , MnO, CaS, MgO is considered to be tapped at $T_M^o K$. Slag heat demand,

$$\Delta H_{slag} = \Delta H_{slag}^S + \Delta H_{slag}^f \quad (3.126)$$

Where, ΔH_{slag}^S is sensible heat of the slag and ΔH_{slag}^f is the heat of slag formation.

(i) Slag constituents CaO, MnO, fractions of SiO_2 , Al_2O_3 , S, MgO comes from reduction

shaft furnace get heated up from $1200^\circ K$ to $T_M^\circ K$ in melter gasifier whereas remaining SiO_2 , Al_2O_3 , S and MgO comes from coal get heated up from $298^\circ K$ to $T_M^\circ K$.

So, sensible heat of slag is,

$$\begin{aligned}\Delta H_{slag}^S = & n_{CaCO_3}^{flux} [55.5 [T_M - 1200]] + n_{CaO}^{ash} [55.5 T_M - 21200] \\ & + n_{SiO_2}^b [72.8 [T_M - 1200]] + n_{SiO_2}^d [72.8 T_M - 26200] \\ & + n_{SiO_2}^g [72.8 [T_M - 1200]] + n_{Al_2O_3}^{ore} [132 [T_M - 1200]] \\ & + n_{Al_2O_3}^{ash} [132 T_M - 55000] + n_{Al_2O_3}^{flux} [132 [T_M - 1200]] \\ & + n_{MnO}^{slag} [57.1 [T_M - 1200]] + n_{CaS}^b [18.7 T_M - 22440] \\ & + n_{CaS}^d [18.7 T_M - 6250] + n_{MgCO_3}^{flux} [55.0 [T_M - 1200]] \\ & + n_{MgO}^{ash} [55.0 T_M - 22300]\end{aligned}\quad (3.127)$$

Where,

$$n_{CaCO_3}^{flux} = n_{CaCO_3}^L + n_{CaCO_3}^D \quad (3.128)$$

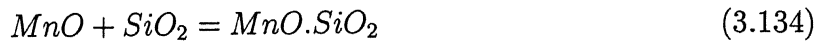
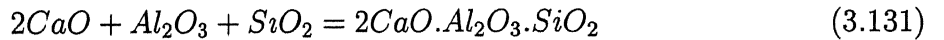
$$n_{Al_2O_3}^{flux} = n_{Al_2O_3}^L + n_{Al_2O_3}^D \quad (3.129)$$

and,

$$n_{MgCO_3}^{flux} = n_{MgCO_3}^L + n_{MgCO_3}^D \quad (3.130)$$

The calculations of $n_{SiO_2}^b$, $n_{SiO_2}^d$, $n_{SiO_2}^g$, n_{CaS}^b and n_{CaS}^d are given in appendix V.

(ii) Following reactions are considered for slag formation :



$$Ca + S = CaS \quad (3.136)$$

So, heat of slag formation,

$$\begin{aligned} \Delta H_{slag}^f = & - \left[2n_{SiO_2}^{slag} - 2n_{MnO}^{slag} - n_{MgO}^{slag} - n_{CaO}^{slag} \right] .90300 \\ & - \left[n_{CaO}^{slag} - n_{Al_2O_3}^{slag} - n_{SiO_2}^{slag} - n_{MnO}^{slag} - 0.5n_{MgO}^{slag} \right] .126840 \\ & - n_{Al_2O_3}^{slag} .104580 - n_{MnO}^{slag} .24780 - 2n_{MgO}^{slag} .63420 \end{aligned} \quad (3.137)$$

Calculations of ΔH_{slag}^f is shown in appendix VI.

So, combining equations (3.126) - (3.137), slag heat demand,

$$\begin{aligned} \Delta H_{slag} = & \left[n_{CaCO_3}^L + n_{CaCO_3}^D \right] [55.5 [T_M - 1200]] + n_{CaO}^{ash} [55.5 T_M - 21200] \\ & + n_{SiO_2}^b [72.8 [T_M - 1200]] + n_{SiO_2}^d [72.8 T_M - 26200] \\ & + n_{SiO_2}^g [72.8 [T_M - 1200]] + n_{Al_2O_3}^{ore} [132 [T_M - 1200]] \\ & + n_{Al_2O_3}^{ash} [132 T_M - 55000] + \left[n_{Al_2O_3}^L + n_{Al_2O_3}^D \right] [132 [T_M - 1200]] \\ & + n_{MnO}^{slag} [57.1 [T_M - 1200]] + n_{CaS}^b [18.7 T_M - 22440] \\ & + n_{CaS}^d [18.7 T_M - 6250] + \left[n_{MgCO_3}^L + n_{MgCO_3}^D \right] [55.0 [T_M - 1200]] \\ & + n_{MgO}^{ash} [55.0 T_M - 22300] - \left[2n_{SiO_2}^{slag} - 2n_{MnO}^{slag} - n_{MgO}^{slag} - n_{CaO}^{slag} \right] .90300 \\ & - \left[n_{CaO}^{slag} - n_{Al_2O_3}^{slag} - n_{SiO_2}^{slag} - n_{MnO}^{slag} - 0.5n_{MgO}^{slag} \right] .126840 \\ & - n_{Al_2O_3}^{slag} .104580 - n_{MnO}^{slag} .24780 - 2n_{MgO}^{slag} .63420 \end{aligned} \quad (3.138)$$

(d) Sensible Heat Of The Gas Leaving Melter-Gasifier

Gas produced in the melter-gasifier containing CO, H_2 and N_2 when no post combustion is done whereas gas contains CO, CO_2 , H_2 , H_2O and N_2 when post combustion of CO and H_2 is done. These gases leaves melter-gasifier at temperature $T_R^\circ K$ and carries away some sensible heat.

So, sensible heat of the gas exiting melter-gasifier,

$$\begin{aligned} \Delta H_{gas} = & n_C^A \left[1 - \frac{ppc}{100} \right] [34.2 T_R - 12600] + \frac{n_H^\circ}{2} \left[1 - \frac{ppc}{100} \right] [30.9 T_R - 10300] \\ & + n_C^A \cdot \left[\frac{ppc}{100} \right] [56.3 T_R - 23100] + \frac{n_H^\circ}{2} \cdot \left[\frac{ppc}{100} \right] [43.7 T_R - 17900] \\ & + n_{N_2}^r [32.1 T_R - 10500] \end{aligned} \quad (3.139)$$

(e) Heat For Drying Of Coal

Water evolves when the temperature of moist coal increases.

$$\text{Moist coal} = \text{Dry coal} + H_2O \quad (3.140)$$

Heat required for drying of coal,

$$\Delta H_{coal} = \frac{M_C \times p_M^{coal}}{1800} [44000] \quad (3.141)$$

(f) Latent Heat Of Devolatilization Of Coal

As the temperature of coal increases, devolatilization occurs according to the reaction :



The latent heat of devolatilization is determined to be 626 KJ/Kg.

So, latent heat of devolatilization of coal,

$$\Delta H_D = 626.M_C \quad (3.143)$$

(g) Heat Losses

We have taken this radiative and convective heat losses as a variable term in our model.

Usually these losses are 20000 KJ/Kg mole of product Fe.

So,

$$\text{Heat losses} = Q_{MG} \quad (3.144)$$

So, total heat demand,

$$\begin{aligned} \Delta H_{dem} &= \Delta H_{HM} + \Delta H_E + \Delta H_{slag} + \Delta H_{gas} + \Delta H_{coal} + \Delta H_D + Q_{MG} \\ &= \left(\frac{100 - X}{95} \right) \times 265000 + \frac{M_C \times p_M^{coal}}{1800} \times [242000] + A \end{aligned} \quad (3.145)$$

Where, A represents all other terms of the equations.

3.3.2 Heat Supply

Heat is supplied in the melter-gasifier by coal combustion and by heat produced for post combustion.

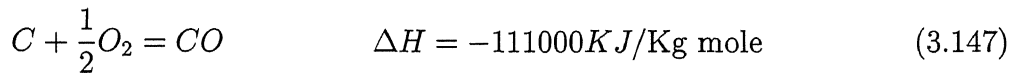
So,

$$\text{Heat supply} = S_{coal} + S_{pc} \quad (3.146)$$

3.3.2.1 Combustion Of Coal

Combustion of coal occurs in the melter-gasifier by pure oxygen blown through the side of the vessel. Combustion of coal produces CO and releases elemental hydrogen.

The reaction occur when coal combusts is,



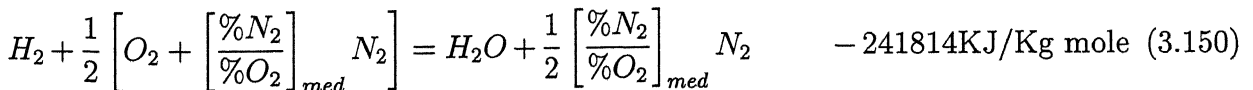
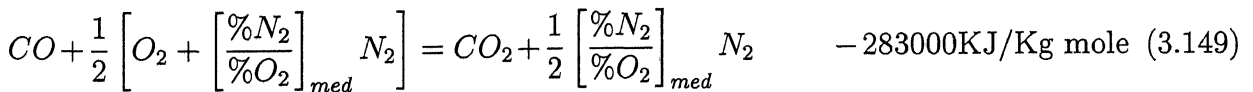
So, heat supplied by coal combustion,

$$S_{coal} = n_C^A [111000] \quad (3.148)$$

3.3.2.2 Post Combustion

The gases produced during the combustion of coal in the melter-gasifier are at or, nearly thermal and chemically equilibrium with the hot metal. The gases produced by combustion of coal are CO, H_2 and N_2 . These CO and H_2 are post combusted by oxygen or, air or, O_2 -enriched air[22]. Necessary amount of oxygen or, air or, O_2 -enriched air are sent from the top of the melter-gasifier.

Following reactions occurs for post combustion :



These two reactions are highly exothermic in nature.

So, produces heat,

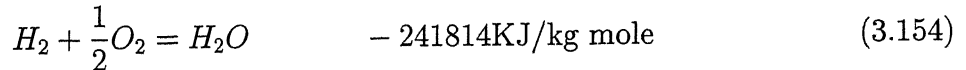
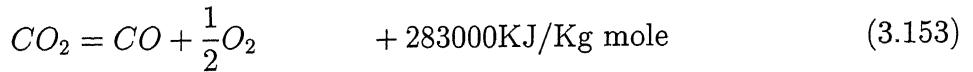
$$S_1 = n_{CO_2}^r \times 283000 + n_{H_2O}^r \times 241814 \quad (3.151)$$

CO_2 and H_2O produced by these two reactions and remaining CO , H_2 and N_2 absorbs all the heat generated and increases gas temperature of melter-gasifier.

Degree of post combustion is defined by[15],

$$ppc = \frac{[\%CO_2 + \%H_2O]}{[\%CO + \%CO_2 + \%H_2 + \%H_2O]} \times 100 \quad (3.152)$$

At higher post combustion ratio, the gas temperature increases which may result dissociation of products of combustion. The following reactions are considered :



Where ϵ is the kg mole of CO_2 dissociated and kg moles of H_2 reacted with oxygen produces by dissociation(Calculation of ϵ is given in appendix VII). So, total heat produced by post combustion,

$$S = (n_{CO_2}^r - \epsilon) \times 283000 + (n_{H_2O}^r + \epsilon) \times 241814 \quad (3.155)$$

Now, heat produced by post combustion is transfered back to the bath. Otherwise, the offgas temperature may be too high which may be problematic for the refractory lining[20].

Fraction heat transfer to the bath[15],

$$HTE = \frac{[\text{heat transfered to bath}] \times 100}{[\text{heat generated by post combustion}]} \quad (3.156)$$

So, heat supplied by post combustion,

$$\begin{aligned} S_{pc} &= S \times \frac{HTE}{100} \\ &= \left[(n_{CO_2}^r - \epsilon) \times 283000 + (n_{H_2O}^r + \epsilon) \times 241814 \right] \times \frac{HTE}{100} \end{aligned} \quad (3.157)$$

Substituting (3.148) and (3.157) into the equation (3.146),

Total heat supply,

$$\Delta H_{sp} = n_C^A \times 111000 + \left[(n_{CO_2}^r - \epsilon) \times 283000 + (n_{H_2O}^r + \epsilon) \times 241814 \right] \times \frac{HTE}{100} \quad (3.158)$$

3.3.3 Output Of The Model

3.3.3.1 Coal Consumption, Flux Rate And Slag Rate

For this calculation, heat balance is done,

$$\text{Heat demand} = \text{Heat supply}$$

Heat demand is taken for the melter-gasifier as the only melter-gasifier heat demand is considered met by coal combustion in our calculation.

So, from equation (3.145) and (3.158),

$$\begin{aligned} \left(\frac{100 - X}{95} \right) \times 265000 + \frac{M_C \times p_m^{coal}}{1800} \times 242000 + A &= n_C^A [111000] \\ &+ \left[(n_{CO_2}^r - \epsilon) \times 283000 + (n_{H_2O}^r + \epsilon) \times 241814 \right] \times \frac{HTE}{100} \end{aligned} \quad (3.159)$$

In this equation X , M_C , M_L , M_D , HTE , ppc , ϵ are unknown. So, if we specify X , ppc , HTE and assume a initial value of M_L , M_D and M_C . Now gas composition and gas temperature can be calculated at that coal amount and flux amount and calculate degree of dissociation(ϵ) of gas component if occurs. H_2O dissociation does not occur within the post combustion range considered. Calculate slag composition and its basicity(B_C). Where,

$$B_C = \frac{n_{CaO}^{slag} + n_{MgO}^{slag}}{n_{SiO_2}^{slag} + n_{Al_2O_3}^{slag}} \quad (3.160)$$

This basicity should match with the required basicity value. After that heat balance is done to check whether heat demand and heat supply are matched or not. So, if these two conditions does not match, then by changing value of M_L , M_D and M_C repeat the process and get actual value of flux rate and coal rate.

So, coal rate(Kg/THM),

$$\begin{aligned} M_C^T &= M_C \cdot \frac{1000}{56} \cdot \frac{p_{Fe}^{HM}}{100} \\ &= M_C \cdot \frac{p_{Fe}^{HM}}{5.6} \end{aligned} \quad (3.161)$$

Similarly, flux rate(Kg/THM),

$$\begin{aligned} M_F^T &= (M_L + M_D) \cdot \frac{1000}{56} \cdot \frac{p_{Fe}^{HM}}{100} \\ &= (M_L + M_D) \cdot \frac{p_{Fe}^{HM}}{5.6} \end{aligned} \quad (3.162)$$

Slag rate are calculated as follows :

Total slag rate per kg moles of product Fe

$$\begin{aligned} M_{slag} &= n_{CaO}^{slag} \times 56.0 + n_{MgO}^{slag} \times 40.0 + n_{Al_2O_3}^{slag} \times 102.0 + n_{MnO}^{slag} \times 70.9 \\ &\quad + n_{CaS}^{slag} \times 72.0 + n_{SiO_2}^{slag} \times 60.0 \end{aligned} \quad (3.163)$$

So, slag rate per ton of hot metal(Kg/THM),

$$\begin{aligned} M_{slag}^T &= M_{slag} \cdot \frac{1000}{56} \cdot \frac{p_{Fe}^{HM}}{100} \\ &= M_{slag} \cdot \frac{p_{Fe}^{HM}}{5.6} \end{aligned} \quad (3.164)$$

3.3.3.2 (CO + H₂) Requirement In The Reduction Shaft Furnace

The gas produced in the melter-gasifier at certain metallization and post combustion containing CO, CO₂, H₂, H₂O, and N₂. The gas contains n_{CO}^r , $n_{CO_2}^r$, $n_{H_2}^r$, $n_{H_2O}^r$ and $n_{N_2}^r$ kg moles CO, CO₂, H₂, H₂O and N₂ respectively. Now the gas is circulated completely

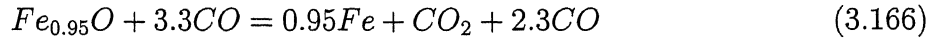
or, partially in the reduction shaft furnace according to the degree of metallization. So, necessary amount of CO and H_2 are sent in the reduction shaft furnace for reduction of iron ore at that metallization.

Now, from Fe-O-C equilibrium curve, it is clear that at $1200^\circ K$ the equilibrium concentration of CO for Fe_3O_4 -wustite is 20% whereas that for wustite to Fe is about 70%. The amount of utilization of CO,

$$\eta_{CO} = 100 \cdot \frac{\%CO_2}{\%CO + \%CO_2} \quad (3.165)$$

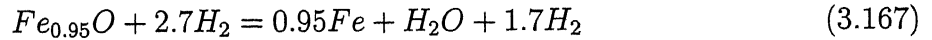
is at a maximum of 80% in respect of magnetite/wustite equilibrium and only 30% for wustite/iron equilibrium.

So, the reaction occurs for wustite reduction is,



From Fe-O-H equilibrium curve, it is also clear that the extent of utilization is 37% for wustite/iron equilibrium.

So, the reaction for wustite reduction by H_2 is,



It is clear from equation (3.166) and (3.167) that one kg mole of H_2 will replace 1.22 kg moles of CO.

So, final equation for reduction of wustite to iron by CO and H_2 takes following form :

$$Fe_{0.95}O + (3.3 - 1.22n_{H_2}^r)CO + n_{H_2}^r H_2 = 0.95Fe + \frac{3.3 - 1.22n_{H_2}^r}{3.3} CO_2 + \frac{2.3}{3.3} [3.3 - 1.22n_{H_2}^r] CO + \frac{n_{H_2}^r}{2.7} H_2O + \frac{1.7}{2.7} n_{H_2}^r H_2 \quad (3.168)$$

H_2 replacement occurs in the range of 0 to 2.7 because when $n_{H_2}^r = 2.7$, the equation (3.168) reduces to (3.167) and when $n_{H_2}^r = 0$, the equation (3.168) reduces to (3.166).

Now, $(CO + H_2)$ generated by this equation (3.169) will be sufficient enough to reduce hematite to magnetite and magnetite to wustite, otherwise iron will not form. So, at certain

metallization ($\text{CO} + \text{H}_2$) left by this reaction will be equal to the requirement of ($\text{CO} + \text{H}_2$) for hematite to wustite reduction. This metallization is called critical metallization.

So, after critical metallization, ($\text{CO} + \text{H}_2$) requirement at X% metallization,

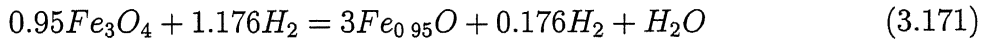
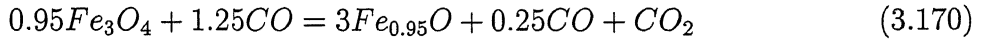
$$n_{\text{COH}_2}^r \text{eq} = \left[\frac{X}{100} \right] \frac{[3.3 - 1.22[n_{\text{H}_2}^r] + [n_{\text{H}_2}^r]]}{0.95} \quad (3.169)$$

Where $[n_{\text{H}_2}^r]$ is kg moles of H_2 in gas at 100% metallization.

Now critical metallization has been calculated as follows :

It is clear from Fe-O-C and Fe-O-H equilibrium curve that the extent of utilization of CO and H_2 for magnetite to wustite reduction are 80% and 85% respectively.

So, these two reactions occurs for magnetite to wustite reduction,



One kg moles of H_2 replaces 1.06 kg moles of CO in this case.

So, final equation for magnetite to wustite reduction by CO and H_2 is,

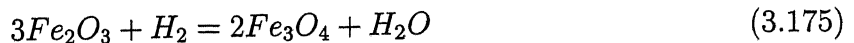
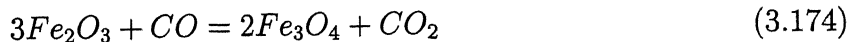
$$\begin{aligned} 0.95\text{Fe}_3\text{O}_4 + (1.25 - 1.06n_{\text{H}_2}^r)\text{CO} + n_{\text{H}_2}^r\text{H}_2 &= 3\text{Fe}_{0.95}\text{O} + \frac{0.25}{1.25}[1.25 - 1.06n_{\text{H}_2}^r]\text{CO} \\ &+ \frac{1.25 - 1.06n_{\text{H}_2}^r}{1.25}\text{CO}_2 + \left[\frac{n_{\text{H}_2}^r}{1.176} \right] \text{H}_2\text{O} + \left[\frac{.176}{1.176} \right] n_{\text{H}_2}^r \cdot \text{H}_2 \end{aligned} \quad (3.172)$$

From equation (3.168), we know that to produce one kg mole product Fe, $\left[\frac{1}{0.95} \right]$ kg moles of $\text{Fe}_{0.95}\text{O}$ is required.

So, to produce $\left[\frac{1}{0.95} \right]$ kg moles of $\text{Fe}_{0.95}\text{O}$ from magnetite(Fe_3O_4) ($\text{CO} + \text{H}_2$) required,

$$= \left[(1.25 - 1.06n_{\text{H}_2}^r) \right] \times \frac{1}{3} \times \frac{1}{0.95} \quad (3.173)$$

For hematite to magnetite reduction these two reactions occurs,



We have to produce $\left[\frac{1}{0.95}\right]$ kg moles of wustite. For that Fe_3O_4 required is $\left[\frac{1}{0.95}\right] \times \left[\frac{0.95}{3}\right] = \frac{1}{3}$ kg moles. (from equation (3.172))

So, to produce $\frac{1}{3}$ kg moles of Fe_3O_4 , kg moles of $(CO + H_2)$ required,

$$= \frac{1}{3} \times \frac{1}{2} = \frac{1}{6} \quad (3.176)$$

So, $(CO + H_2)$ required for hematite to wustite reduction,

$$= [(1.25 - 1.06n_{H_2}^r) + n_{H_2}^r] \times \frac{1}{3} \times \frac{1}{0.95} + \frac{1}{6} \quad (3.177)$$

From equation (3.168), $(CO + H_2)$ left,

$$= \left[\frac{X}{100}\right] \left[\frac{2.3}{3.3}(3.3 - 1.22n_{H_2}^r) + \frac{1.7}{2.7}n_{H_2}^r\right] \quad (3.178)$$

At critical metallization, this $(CO + H_2)$ left must be equal to $(CO + H_2)$ requirement for hematite to wustite reduction in the reduction shaft furnace.

So, by equating (3.177) and (3.178), we calculate the value of critical metallization.

3.3.3.3 Oxygen Requirement, Top Gas Amount And Composition

Coal rate are already calculated at a certain metallization and post combustion ratio. Now, we know the value of M_L and M_D also. So, by knowing the gas composition (n_{CO}^r , $n_{CO_2}^r$, $n_{H_2}^r$, $n_{H_2O}^r$, $n_{N_2}^r$) oxygen required for post combustion have been calculated.

$$\begin{aligned} \text{Oxygen required for post combustion}(Nm^3/THM) &= n_{O_2}^{pc} \times 22.4 \times \frac{p_{Fe}^{HM}}{5.6} \\ &= \frac{1}{2}(n_{CO_2}^r + n_{H_2O}^r) \times 22.4 \times \frac{p_{Fe}^{HM}}{5.6} \end{aligned} \quad (3.179)$$

$$\text{As, } n_O^{pc} = n_{CO_2}^r + n_{H_2O}^r$$

So, O_2 required for coal combustion have been calculated by O-balance equation. (n_O^B)

So, O_2 required for coal combustion (Nm^3/THM),

$$= \frac{1}{2}n_O^B \times 22.4 \times \frac{p_{Fe}^{HM}}{5.6} \quad (3.180)$$

Amount of top gas component in Nm^3/THM are determined.

$$\text{CO in the top gas, } (CO^{tg}) = n_{CO}^{tg} \times 22.4 \times \frac{p_{Fe}^{HM}}{5.6}$$

Similarly CO_2^{tg} , H_2^{tg} , H_2O^{tg} and N_2^{tg} have been calculated. The calculation of n_{CO}^{tg} , $n_{H_2}^{tg}$, $n_{H_2O}^{tg}$, $n_{CO_2}^{tg}$ and $n_{N_2}^{tg}$ are shown in appendix III.

So, total top gas amount(Nm^3/THM)

$$= CO^{tg} + CO_2^{tg} + H_2^{tg} + H_2O^{tg} + N_2^{tg} \quad (3.181)$$

3.3.3.4 Export Gas

The export gas is calculated by considering top gas which is coming out from reduction shaft furnace and surplus gas which is coming out from melter-gasifier. Surplus gas is calculated by considering production of the gas in the melter-gasifier and requirement of the reduction gas in the reduction shaft furnace for the reduction of iron ore. At higher degree of metallization the requirement of the gas in the RSF is more than at lower metallization. So, surplus gas is calculated by :

$$\text{Surplus gas } [n_t^{sg}] = \text{Production} - \text{Requirement} \quad (3.182)$$

Top gas comes out from reduction shaft furnace is cooled and cleaned in a scrubber. After which it is available as a highly valuable export gas.

Total export gas = Surplus gas + Top gas

So,

$$n_t^{ex} = n_t^{sg} + n_t^{tg} \quad (3.183)$$

Export gas is suitable for a wide range of application including power generation, production of DRI, synthetic gas generation etc.

The calculation of n_t^{sg} and n_t^{tg} are given in appendix III.

3.4 Method Of Calculation

For this thermochemical model, a computer program is developed to calculate coal rate, flux rate, slag rate, oxygen requirement and export gas as a function of degree of metallization,

post combustion ratio, heat transfer efficiency, coal rank. Figure 3.2 shows the flowchart of the computer program. For a guess value of flux and coal amount first it calculates kg moles of materials comes in and goes out. Then solve material balance to get gas composition in the melter-gasifier. Gas temperature equation(f) and the dissociation equation(m) are solved to get the gas temperature and degree of dissociation. Calculate slag composition at this amount of coal rate and flux rate. Basicity(B_C) is calculated from this slag composition. After this heat balance equations are solved. If heat demand and heat supply terms are not near about equal and basicity(B_C) does not match with the given basicity(B) value. Then change the value of flux and coal amount and repeat this process until both conditions are fulfilled. When it matches, then calculate coal rate and flux rate. After knowing this coal rate and flux rate, oxygen requirement, slag rate, ore rate, gas temperature, dissociation are calculated. Requirement of the reduction gas in the reduction shaft furnace are calculated. From this requirement top gas amount and surplus gas are calculated. Finally adding this top gas and surplus gas, export gas has been calculated.

3.5 Validation Of The Model

Delpont has published the data of a commercial scale Corex plant on consumption of coal, ore, flux, slag and oxygen for several compositions of hot metal and different tapping temperatures of hot metal and slag[13]. In the reduction shaft furnace, they charged iron ore and a mixture of limestone and dolomite which were metallized and preheated to $1200^\circ K$ respectively by the hot reduction gases supplied from the melter-gasifier. The composition of iron ore was :

Sishen ore : 66.3%Fe, 0.05%Mn, 0.03%P, 0.02%S, 3.3% SiO_2 , 1.2% Al_2O_3

Thabazimbi ore : 62.9%Fe, 0.10%Mn, 0.04%P, 0.005%S, 5.2% SiO_2 , 1.2% Al_2O_3

CVRD ore : 65.1%Fe, 0.02%Mn, 0.009%P, 0.004%S, 4.7% SiO_2 , 0.6% Al_2O_3

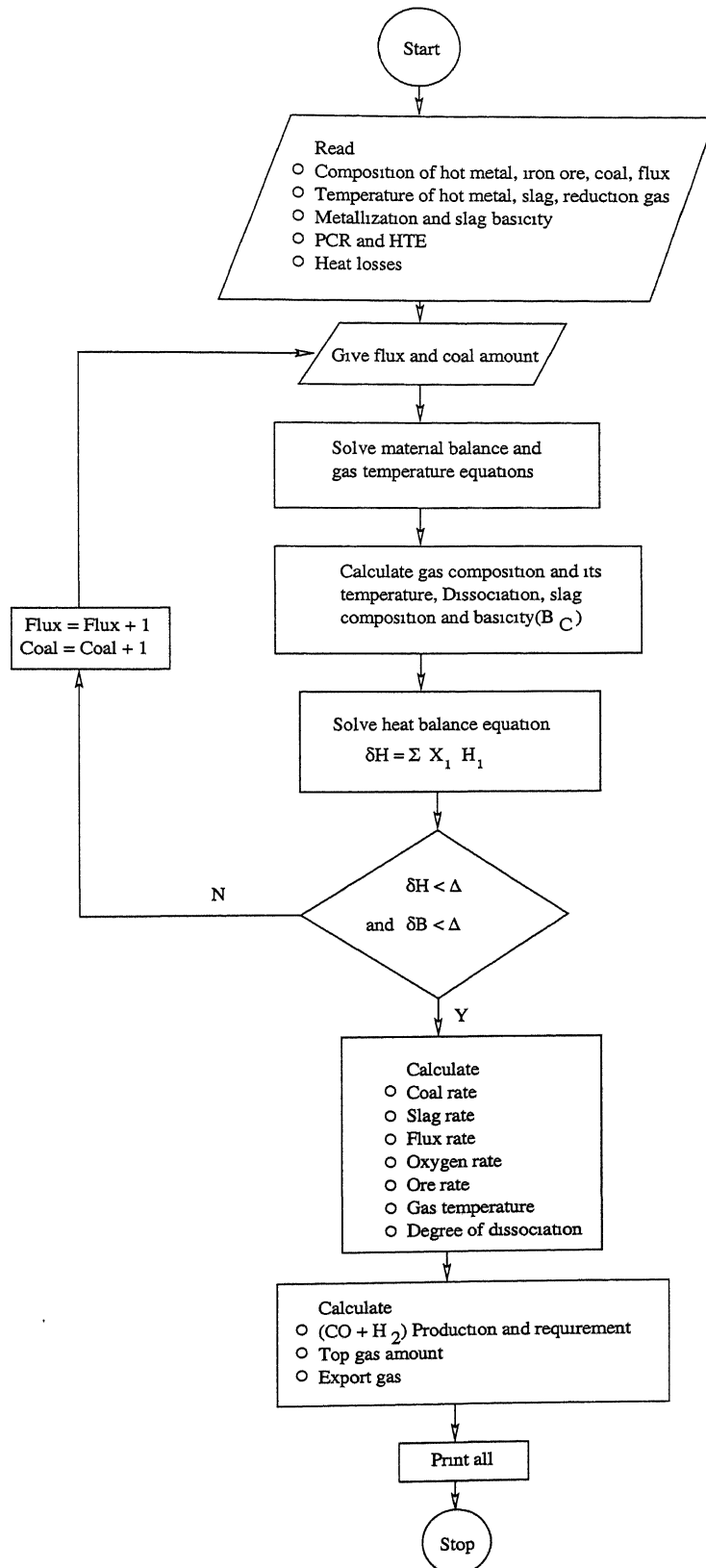


Figure 3.2: Flowchart Of Computer Program

The composition of coal was :

Ash = 16.24%, Volatiles = 27.43%, C = 80.59%, H = 4.86%, O = 12.33%, N = 1.81%, S = 0.87%, P = 0.015%

The composition of ash was :

SiO_2 = 41.22%, Al_2O_3 = 25.62%, CaO = 10.19%, MgO = 2.17%

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Coal was charged into the melter-gasifier and the hot reduction gas at $1273^\circ K$ was circulated into the reduction shaft furnace. All the reported data on the composition of hot metal, coal, ore and temperature of hot metal and slag are used to compute the consumption of coal, ore, flux, slag and oxygen rate per ton of hot metal by the model of present study. Some of the data on the degree of metallization of iron ore, basicity of the slag, proportion of limestone and dolomite and the heat losses occurring are not given and hence these values are assumed. In the calculations, it is assumed that the slag basicity is 1.2 and is obtained by the addition of equal proportion of limestone (96.5%CaCO₃, 1.8%MgCO₃, 0.5%SiO₂ and 0.5%Al₂O₃) and dolomite(54.6%CaCO₃, 43.8%MgCO₃, 0.5%SiO₂ and 0.5%Al₂O₃). The selected values of basicity and flux composition are commonly used in the ironmaking technology[10]. According to Delport, a high(> 90%) and stable degree of metallization of iron ore is necessary to maintain good hot metal quality[13]. In the calculations 95% metallization is taken.

Figure 3.3 to 3.7 compare the calculated values of coal rate, oxygen rate, flux rate, slag rate and ore rate with the reported ones for different compositions of hot metal and temperature of hot metal and slag and values are given in table 3.1.

Though the calculated values of all are close to the reported ones but the exact match between the two values could not be obtained.

This discrepancy in all the computed values is most probably due to nonavailability of the actual value of the metallization of iron ore, basicity of slag, proportion of limestone and dolomite and the heat losses. It has been observed that a four percent change in metallization changes the coal rate by 6%, slag rate by 8% and oxygen rate by 13%. Similarly ten percent

change in basicity changes coal rate by 10%, flux rate by 13%, oxygen rate by 22%. Twenty percent change in heat losses changes the value of all the outputs by 4-5%. Thus it appears that the model predictions on outputs are reasonably in agreement with the reported ones

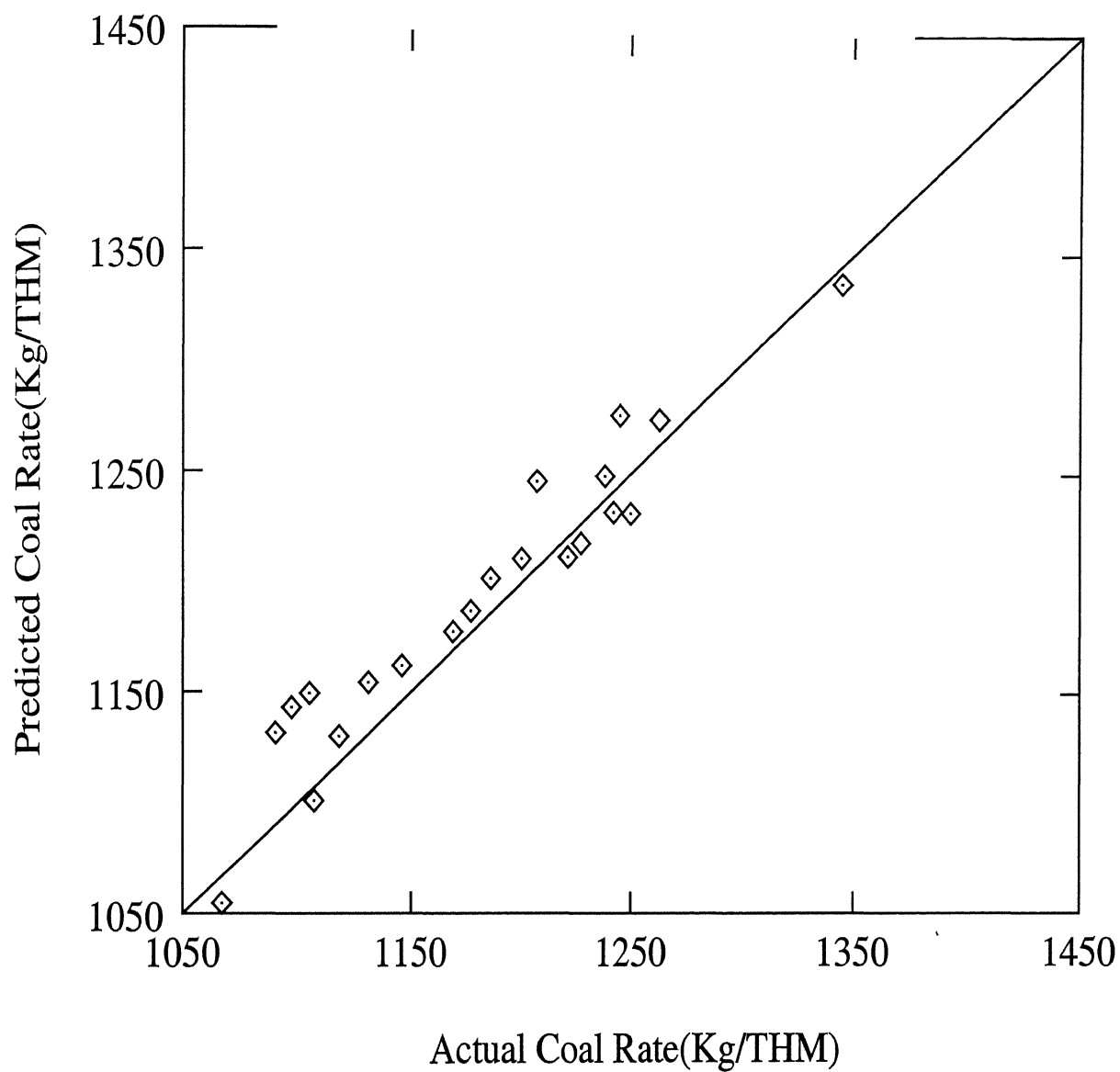


Figure 3.3: Actual coal rate vs Predicted coal rate

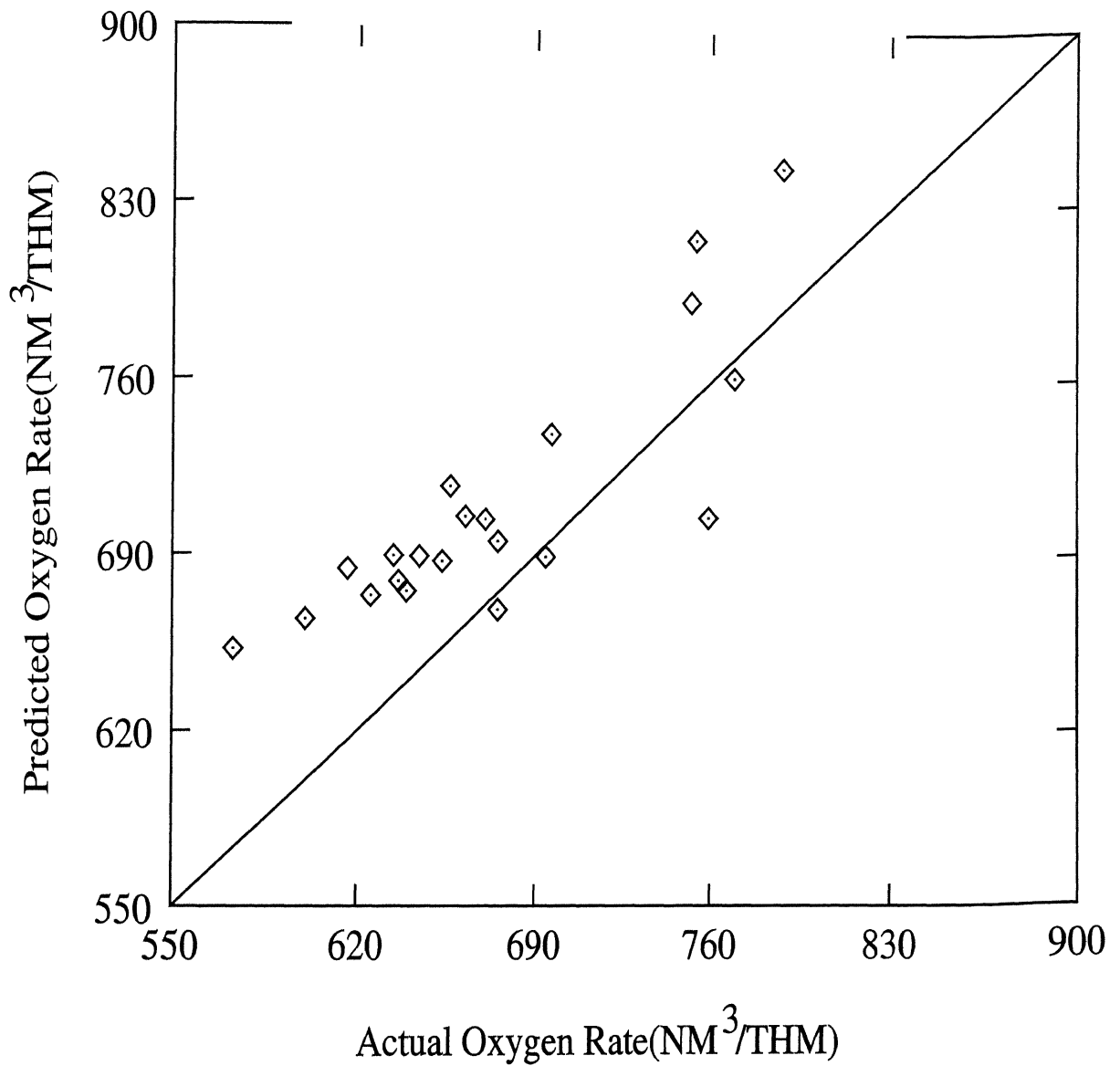


Figure 3.4: Actual oxygen rate vs Predicted oxygen rate

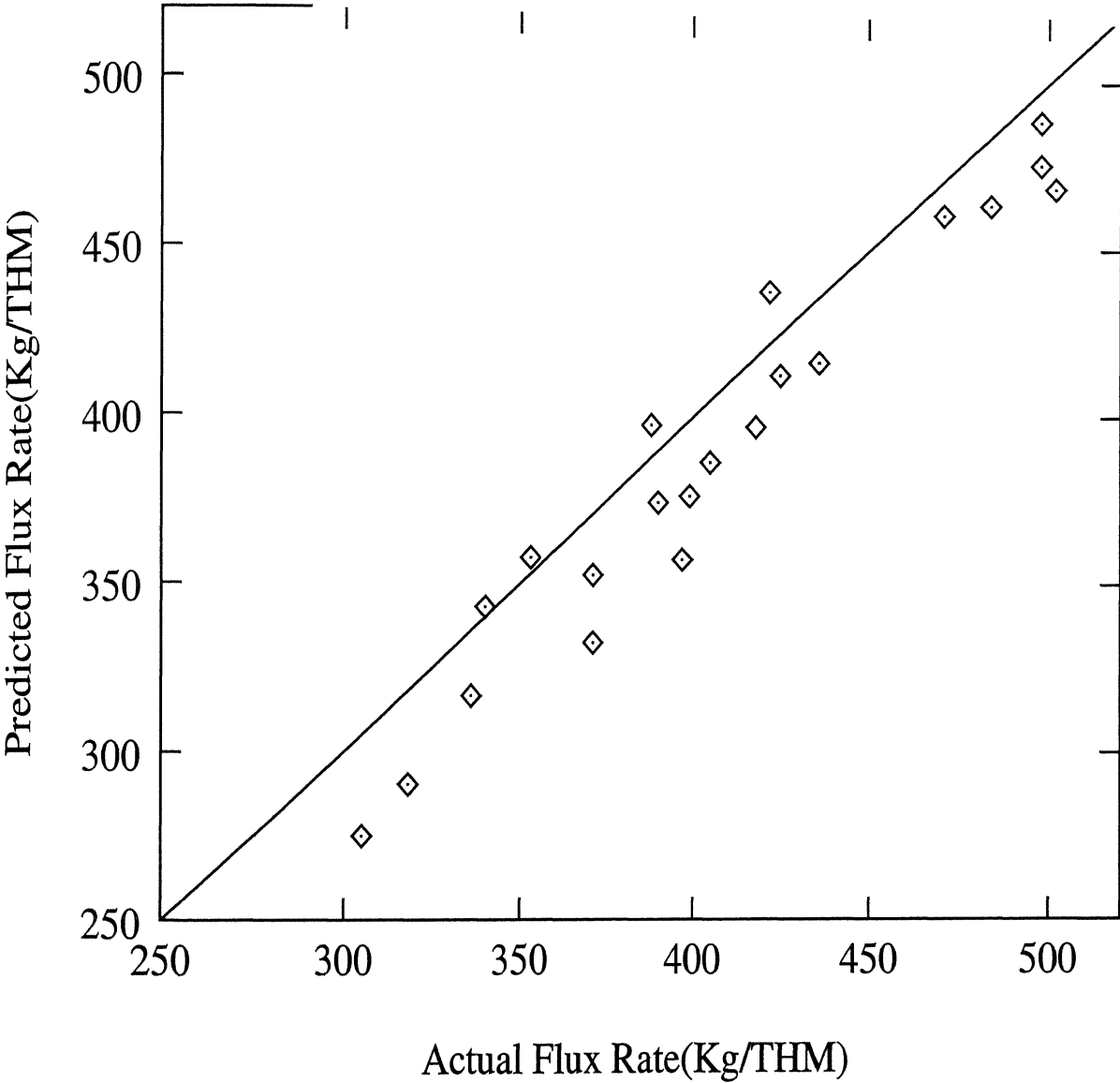


Figure 3.5: Actual flux rate vs Predicted flux rate

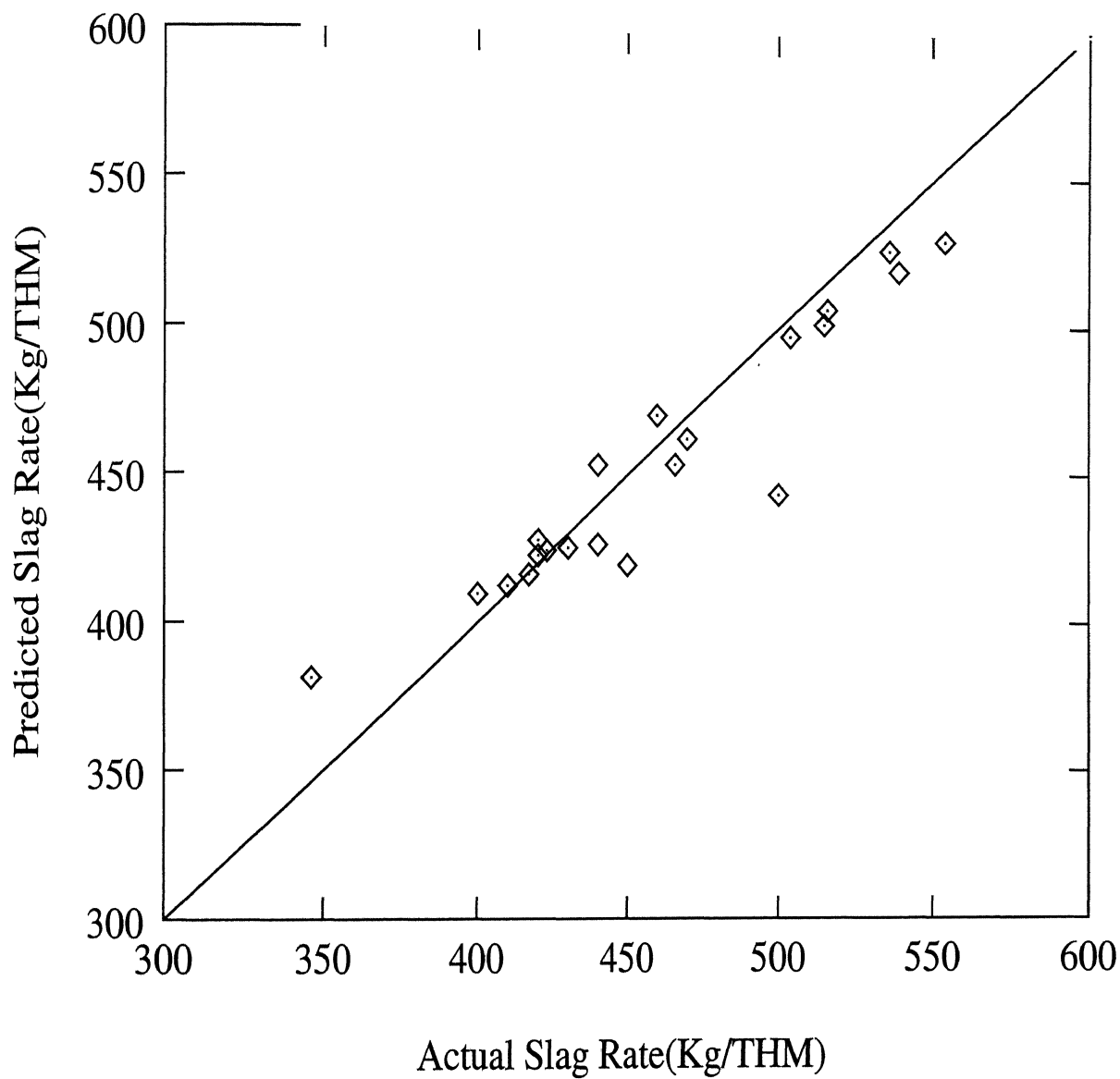


Figure 3.6: Actual slag rate vs Predicted slag rate

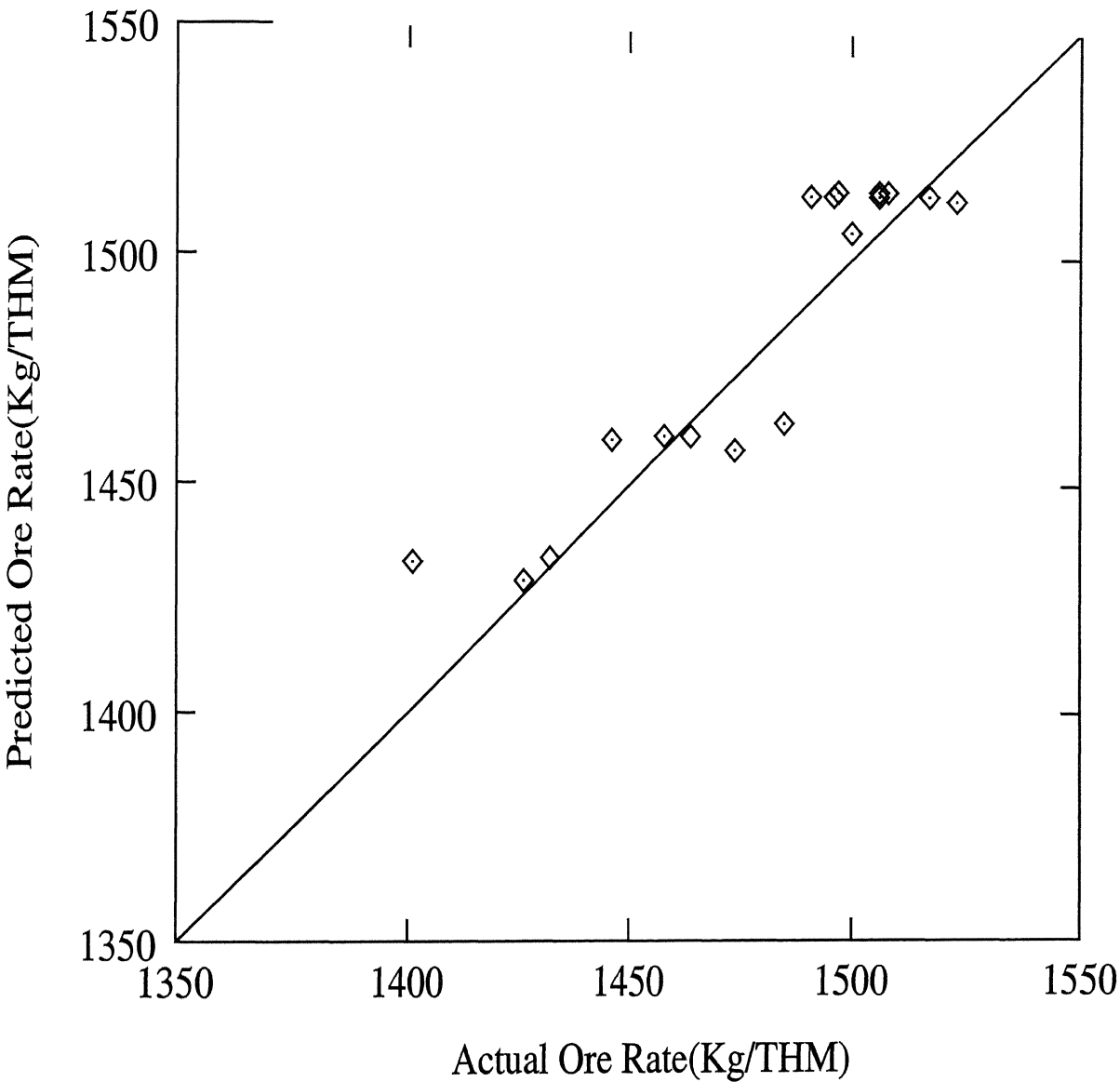


Figure 3.7: Actual ore rate vs Predicted ore rate

Chapter 4

Results And Discussions

A thermochemical model results coal consumption, oxygen requirement, flux rate, slag rate and export gas. The results of the calculations are in the following figures presents for the following fixed compositions of input :

Hot metal composition :

$$\%C = 4.06 \quad \%Si = 0.39 \quad \%Mn = 0.6 \quad \%S = 0.109 \quad \%P = 0.152$$

Ore composition :

$$\%Fe_2O_3 = 94.7 \quad \%SiO_2 = 3.3 \quad \%MnO = .06 \quad \%Al_2O_3 = 1.2 \quad \%P_2O_5 = 0.07$$

$$\%FeS = 0.055$$

Limestone composition :

$$\%CaCO_3 = 96.5 \quad \%MgCO_3 = 1.8 \quad \%Al_2O_3 = 0.5 \quad \%SiO_2 = 0.5$$

Dolomite composition :

$$\%CaCO_3 = 54.6 \quad \%MgCO_3 = 49.8 \quad \%Al_2O_3 = 0.5 \quad \%SiO_2 = 0.5$$

Limestone : Dolomite = 50 : 50

Ash composition :

$$\%SiO_2 = 61.66 \quad \%Al_2O_3 = 38.34$$

Hot metal temperature = $1527^{\circ}C$

(Flux + Metallized ore) temerature = $1200^{\circ}K$

Top gas temperature = $200^{\circ}C$

Heat losses = 20000 KJ/Kg mole of product Fe

Calculations are done for these :

Degree of metallization = (0 - 100%)

Post combustion ratio = (0 - 50%)

Heat transfer efficiency = (70 - 100%)

Medium of post combustion are O_2 , air and O_2 -enriched air.

Coal rank :

Bituminus, subbituminus and anthracite.

(Compositions are given in appendix VIII).

Basicity of slag = 1.2

The calculations are made by calculating heat demand and supply in the system. The melter-gasifier heat demand is met by coal combustion.

In figure 4.1 heat demand for melter-gasifier has been plotted against degree of metallization. In this figure heat demand for melter-gasifier are drawn for both hot metal and slag temperature of $1527^\circ C$ and $1427^\circ C$. Line (a) and (b) shows melter-gasifier heat demand when hot metal and slag temperatures are $1527^\circ C$ and $1427^\circ C$ respectively. The heat demand for melter-gasifier decreases gradually. When hot metal and slag are tapped at $1527^\circ C$, then at 0% metallization heat demand for melter-gasifier is 16.6 GJ/thm and it decreases to 6.8 GJ/thm at 100% metallization. Because as the metallization increases, less reduction of pre-reduced iron ore is done in the melter-gasifier. At 100% metallization, as there are no reduction of iron ore in the melter-gasifier, here heat demand is minimum. Similarly when hot metal and slag both are tapped at $1427^\circ C$, then heat demand for melter-gasifier varies from 14.8 GJ/thm to 6.1 GJ/thm when metallization varies from 0% to 100% shown in line (b) in this figure.

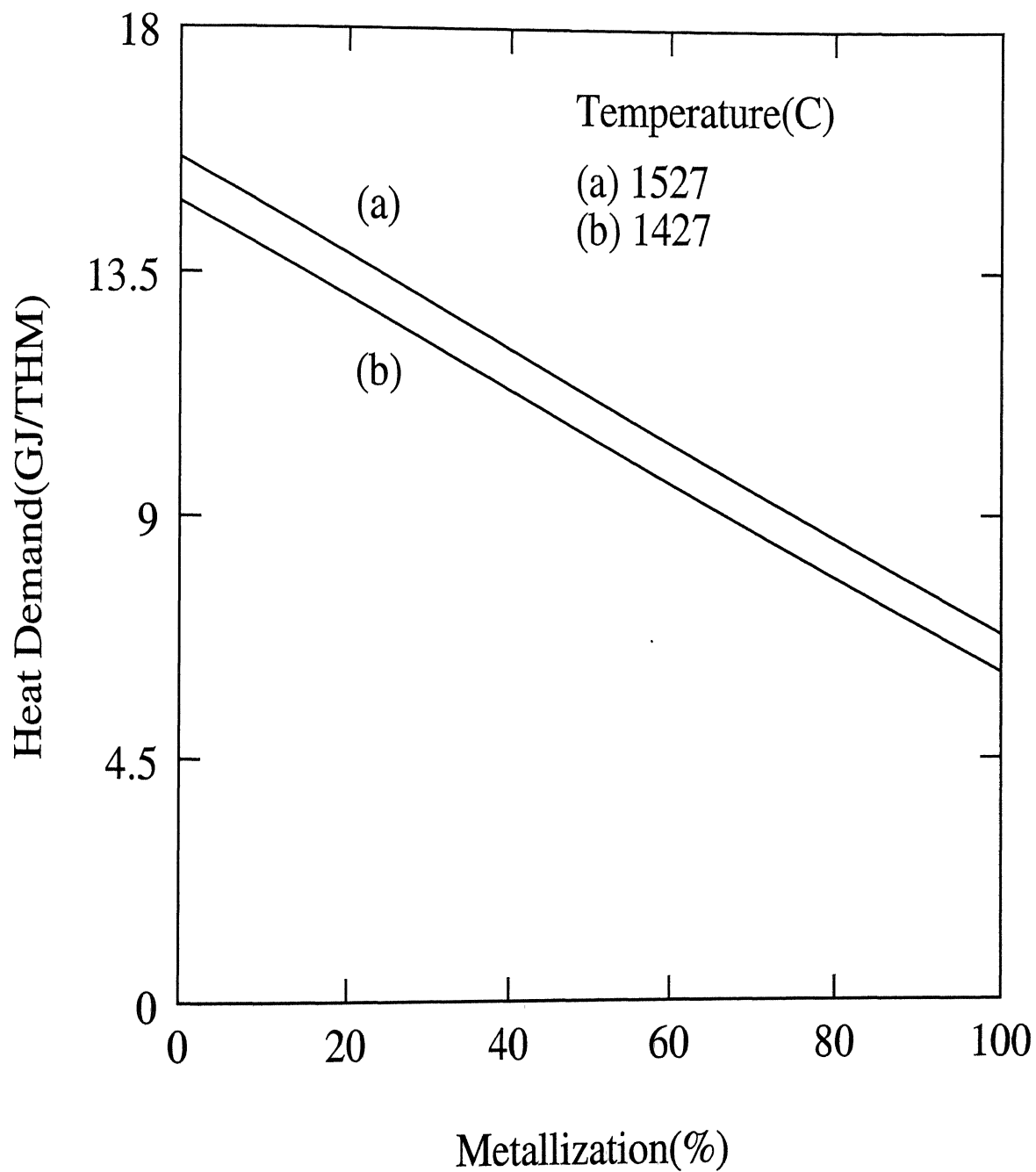


Figure 4.1: Heat demand vs degree of metallization

Heat demand in the melter-gasifier for different metallurgical operations are given in table 4.1.

4.1 Effect of Degree of Metallization

The effect of degree of metallization of iron ore on coal rate, O_2 -requirement, flux rate, slag rate and export gas are discussed in this section. In this presentation, following are considered :

Ultimate analysis of bituminous coal : %C = 86.4, %H = 4.9, %O = 3.6, %N = 1.6, %S = 0.6 and % ash = 2.9 ; Proximate analysis : %VM = 22.8, % Fixed carbon = 77.2

Ultimate analysis of subbituminous coal : %C = 64.2, %H = 4.4, %O = 16.6, %N = 1.2, %S = 0.4 and % ash = 13.2 ; Proximate analysis : %VM = 44.4, % Fixed carbon = 54.6

Ultimate analysis of anthracite coal : %C = 84.2, %H = 2.8, %O = 2.2, %N = 0.8, %S = 0.6 and % ash = 9.4 ; Proximate analysis : %VM = 3.7, % Fixed carbon = 96.3

Post combustion ratio = 0%

4.1.1 Coal Rate

Consumption of coal is the important technoeconomic index of the COREX process. In figure 4.2 coal consumption has been plotted against degree of metallization for different coal grade. Figure shows that as the degree of metallization increases, coal consumption decreases. Because as the degree of metallization increases, heat demand in the melter-gasifier decreases shown in figure 4.1. Coal rate at various degrees of metallization for different coal grade are given in table 4.2.

For bituminous coal, the coal consumption decreases from 1977.24 Kg/thm to 750.61 Kg/thm as the degree of metallization increases from 0% to 100%. For subbituminous coal it is decreases from 2440.1 Kg/thm to 927.9 Kg/thm. Figure shows that subbituminous coal consumption is more than bituminous and anthracite coal. Because normally

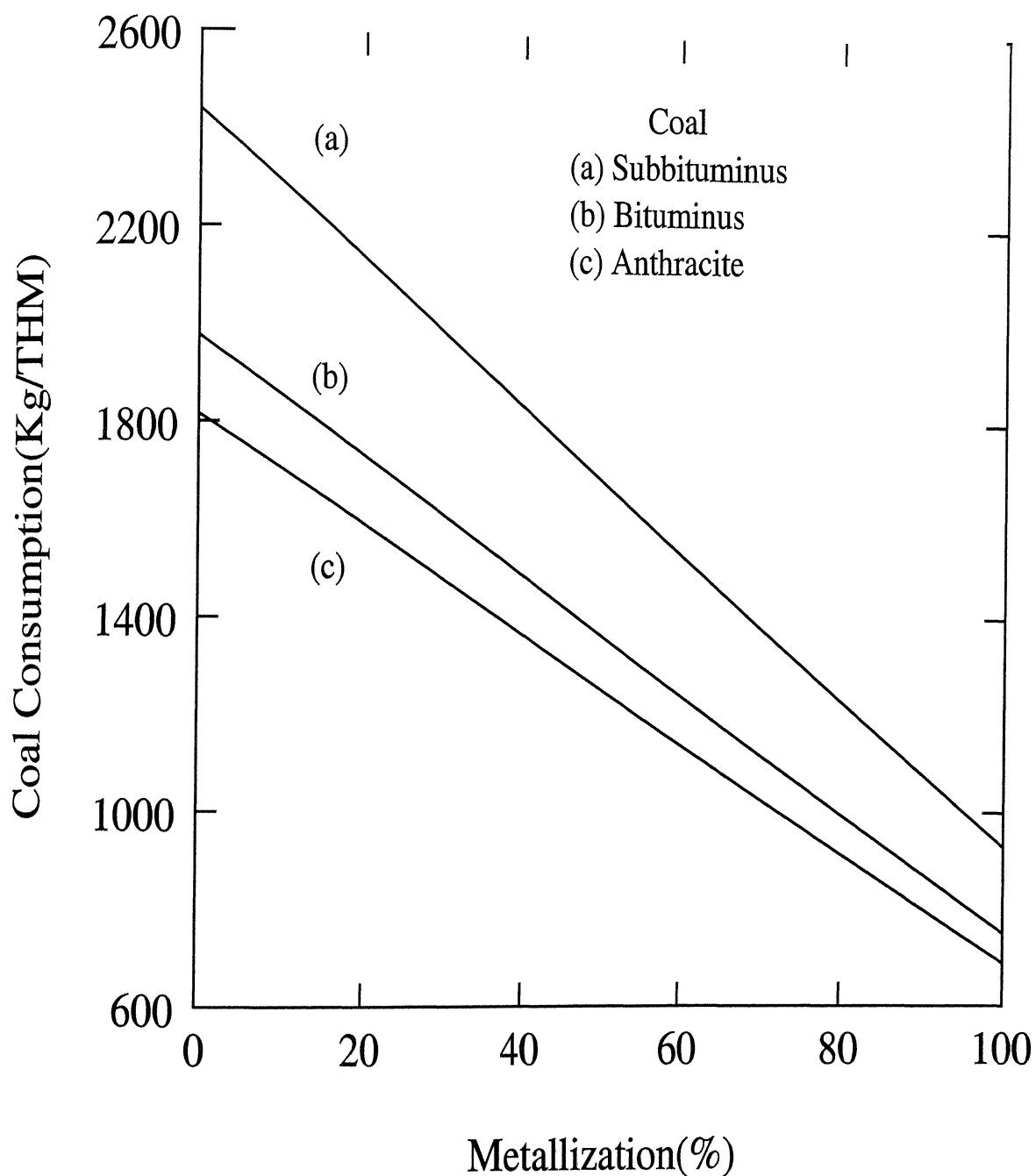


Figure 4.2: Coal consumption vs degree of metallization for various types of coal

coal is characterized by its volatile matter. As volatile matter content in subbituminous coal is highest i.e. 44.4%, fixed carbon is 54.6% lowest. For anthracite coal volatile matter content is less i.e. 3.7% and fixed carbon is 96.3%. As the fixed carbon content is lowest in subbituminous coal, more coal is consumed. But in anthracite coal, as fixed carbon is more, coal consumption is less. Similarly bituminous coal consumption is lower than subbituminous coal and higher than anthracite coal as the fixed carbon content in the bituminous coal is in between subbituminous and anthracite coal.

4.1.2 Oxygen Requirement

Oxygen requirement has been plotted against degree of metallization for various types of coal in figure 4.3. Oxygen requirement decreases with degree of metallization increases because oxygen requirement is directly proportional to coal consumption. As the degree of metallization increases, coal consumption decreases, hence oxygen requirement for coal combustion decreases. Oxygen requirement decreases from $3444.9 \text{ Nm}^3/\text{thm}$ to $1077 \text{ Nm}^3/\text{thm}$ when bituminous coal is used for coal combustion as the degree of metallization increases from 0% to 100%. When subbituminous coal is used, oxygen requirement is more than oxygen requirement when bituminous coal is used because subbituminous coal consumption is more than bituminous coal. Oxygen requirement for subbituminous coal is decreases from $4241.6 \text{ Nm}^3/\text{thm}$ at 0% metallization to $1319 \text{ Nm}^3/\text{thm}$ at 100% metallization (see table 4.5). Similar trend is seen for anthracite coal.

4.1.3 Flux and Slag Rate

Flux rate decreases with degree of metallization shown in figure 4.4 for various types of coal. In this figure we see that at 0% metallization flux rate is 284 Kg/thm , but it decreases to 212.7 Kg/thm at 100% metallization when bituminous coal is used. Flux rate decreases with degree of metallization because flux contains limestone and dolomite and both contains CaCO_3 and MgCO_3 from where most of CaO and MgO comes in the slag. As the coal

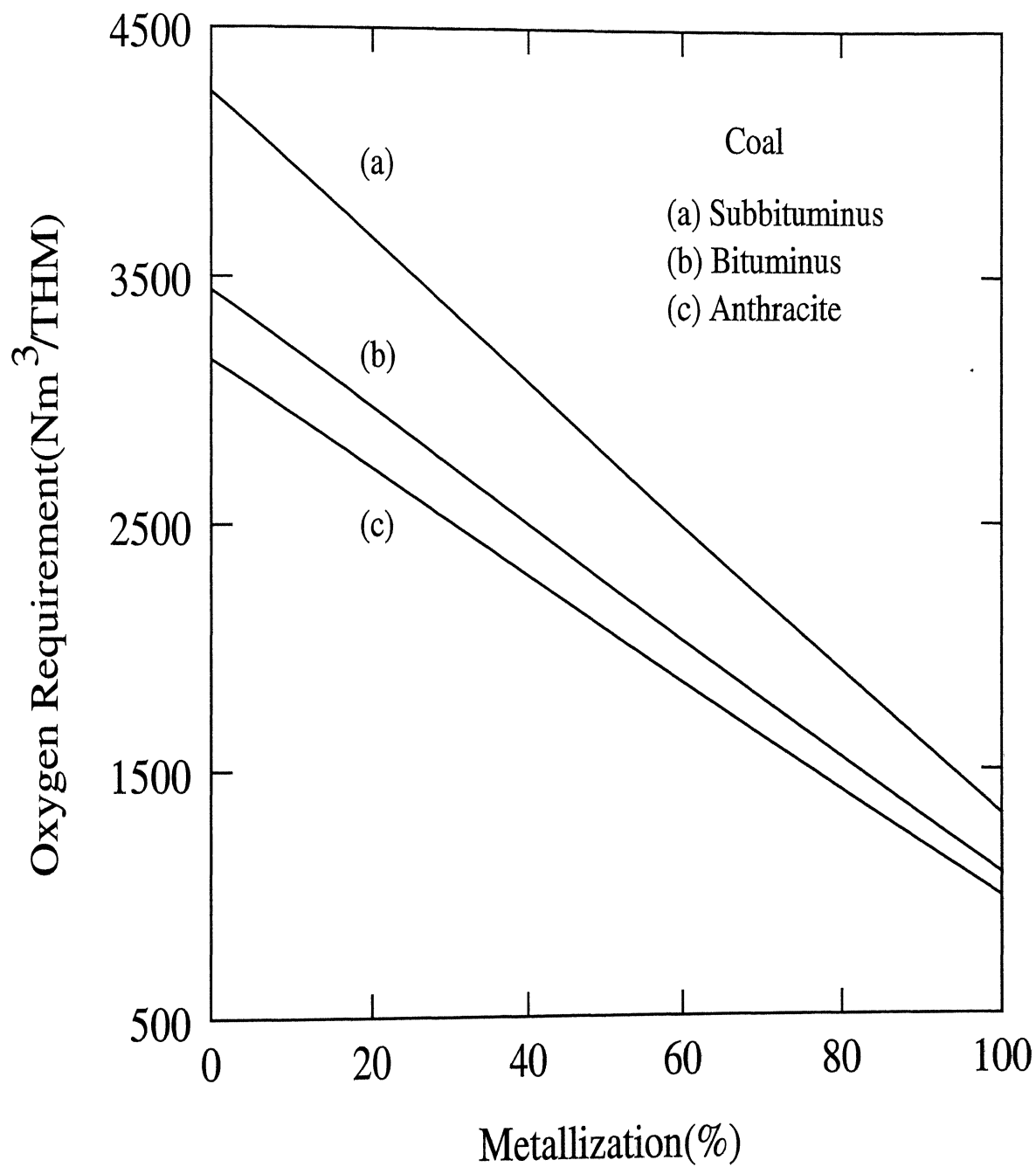


Figure 4.3: Oxygen requirement vs degree of metallization for various types of coal

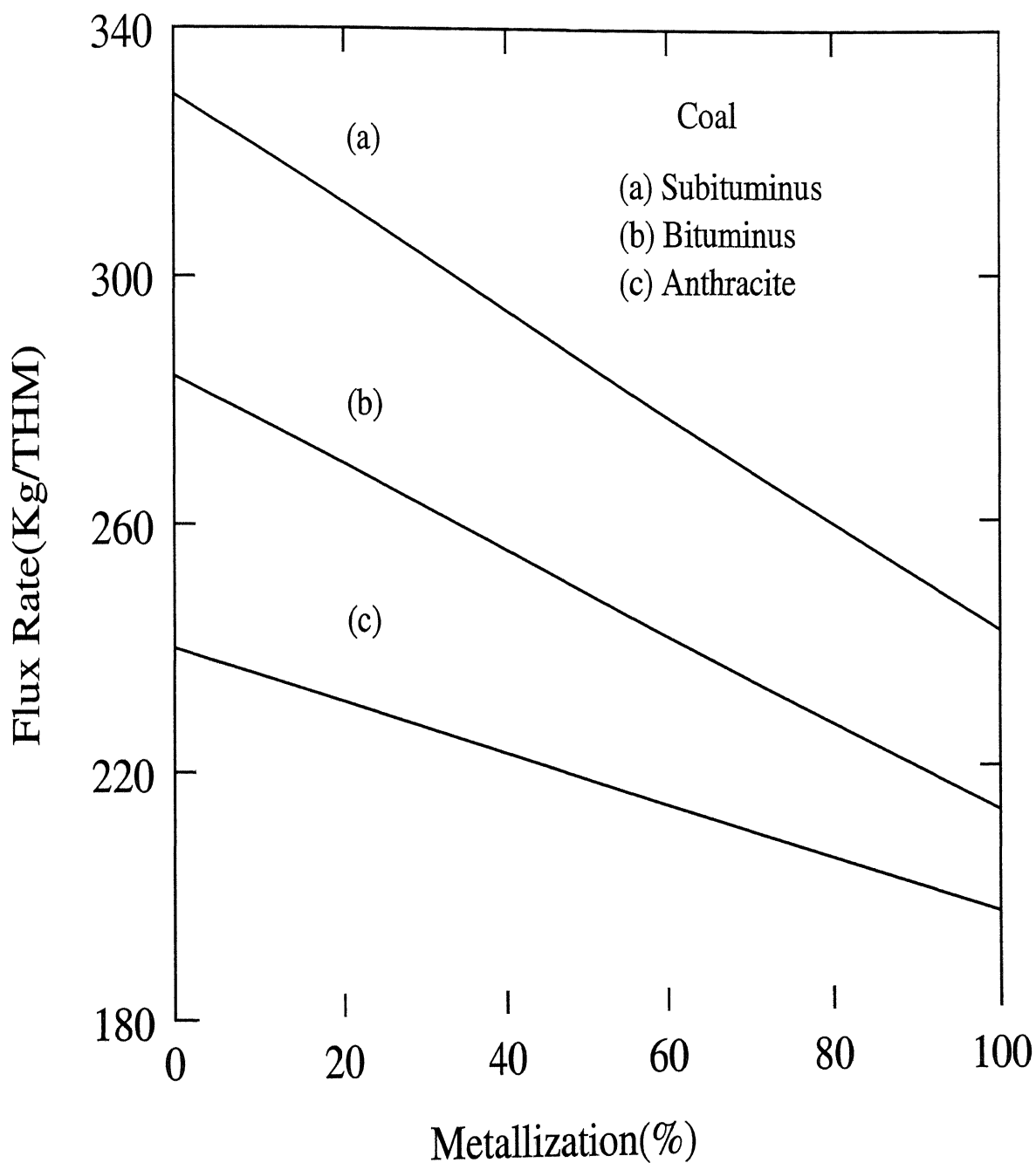


Figure 4.4: Flux rate vs degree of metallization for various types of coal

consumption decreases with degree of metallization, so SiO_2 and Al_2O_3 in the slag coming less at higher degree of metallization because SiO_2 and Al_2O_3 comes from iron ore and coal. As the iron ore amount is fixed in any degree of metallization, SiO_2 and Al_2O_3 comes from coal is less at high degree of metallization. Basicity is the ratio of $(CaO + MgO)$ and $(SiO_2 + Al_2O_3)$ in the slag. As the basicity is fixed i.e. 1.2. So, to maintain this basicity of slag, flux rate decreases with degree of metallization. Similarly when subbituminous and anthracite coal are used, flux requirement decreases with degree of metallization also shown in this figure. Flux rate at various degree of metallization is given in table 4.8.

Slag rate is proportional to flux rate and coal rate. As both flux rate and slag rate are low at higher degree of metallization. So, slag rate also decreases with degree of metallization. At 0% metallization slag rate is 308.7 Kg/thm and decreases to 231.6 Kg/thm when degree of metallization increases to 100% for bituminous coal. For subbituminous coal, slag rate varies from 374.3 Kg/thm to 275.6 Kg/thm when metallization increases from 0% to 100%(see table 4.9). For anthracite coal also is shown in this figure 4.5.

4.1.4 Export Gas

In the melter-gasifier gas produced by coal combustion contains CO , H_2 and N_2 . The gas is cleaned in hot cyclone and fed to reduction shaft furnace as reducing gas according to the requirement of CO and H_2 for reduction of iron ore in the reduction shaft furnace. The rest gas comes out from melter-gasifier known as surplus gas. After reduction of iron ore by CO and H_2 some amount of gas comes out from reduction shaft furnace according to utilization of CO and H_2 known as top gas. Surplus gas is mixed with top gas, known as export gas. This gas is used for many purposes like in the generation of electrical energy, as synthesis gas in chemical industry etc.

So,

$$\text{Export gas} = \text{surplus gas} + \text{top gas}$$

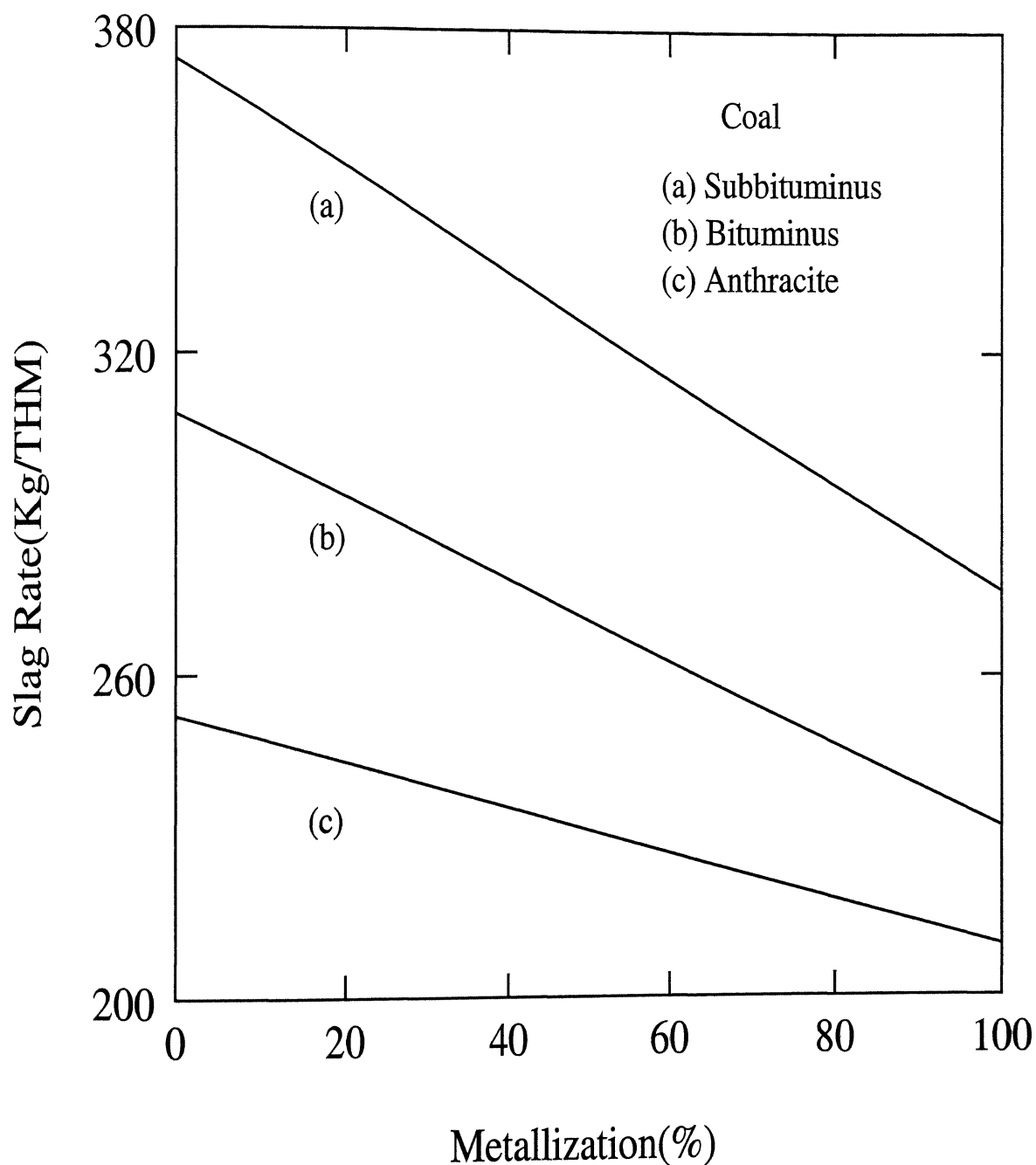


Figure 4.5: Slag rate vs degree of metallization for various types of coal

Now,

$$\text{Surplus gas} = \text{Production of } (\text{CO} + \text{H}_2) \text{ in MG} - \text{Requirement of } (\text{CO} + \text{H}_2) \text{ in RSF.}$$

Production of $(\text{CO} + \text{H}_2)$ in the melter-gasifier and requirement of $(\text{CO} + \text{H}_2)$ in the reduction shaft furnace has been plotted against degree of metallization in figure 4.6 when bituminous coal is used for coal combustion. Production of $(\text{CO} + \text{H}_2)$ in the melter-gasifier decreases with degree of metallization because of less coal consumption. $(\text{CO} + \text{H}_2)$ requirement for reduction of iron ore in the reduction shaft furnace increases with degree of metallization because more reduction has to be carried out in the reduction shaft furnace at higher degree of metallization. $(\text{CO} + \text{H}_2)$ requirement has been calculated from $\text{Fe}_{0.95}\text{O}$ -Fe thermodynamic equilibrium at 1200°K . $(\text{CO} + \text{H}_2)$ left after $\text{Fe}_{0.95}\text{O}$ -Fe reaction must be sufficient enough to carry out Fe_2O_3 - Fe_3O_4 and Fe_3O_4 - $\text{Fe}_{0.95}\text{O}$ reaction upto a certain degree of metallization. This is known as critical metallization is 30% shown in this figure 4.6.

It is clear from figure 4.6 that as the degree of metallization increases, $(\text{CO} + \text{H}_2)$ production in the melter-gasifier decreases and requirement in the reduction shaft furnace increases. So, with more degree of metallization, surplus gas comes out from melter-gasifier is less than at lower degree of metallization. Because here more gas is produced in the melter-gasifier and sent less in the reduction shaft furnace. For this reason surplus gas decreases with degree of metallization shown in figure 4.7 and all values of surplus gas are given in table 4.10. The figure 4.7 is drawn for bituminous coal. Surplus gas decreases from $4058 \text{ Nm}^3/\text{thm}$ to $372.5 \text{ Nm}^3/\text{thm}$ as the degree of metallization increases from 0% to 100%.

Since 30% utilization of CO and 37% utilization of H_2 is considered for $\text{Fe}_{0.95}\text{O}$ -Fe equilibrium. So, large amount of CO and H_2 is obtained in the top gas. As the degree of metallization increases, $(\text{CO} + \text{H}_2)$ in the top gas increases because more $(\text{CO} + \text{H}_2)$ is sent in the reduction shaft furnace according to the requirement for reduction at higher degree of metallization. Top gas $(\text{CO} + \text{H}_2)$ increases from $154.4 \text{ Nm}^3/\text{thm}$ to $831 \text{ Nm}^3/\text{thm}$ when

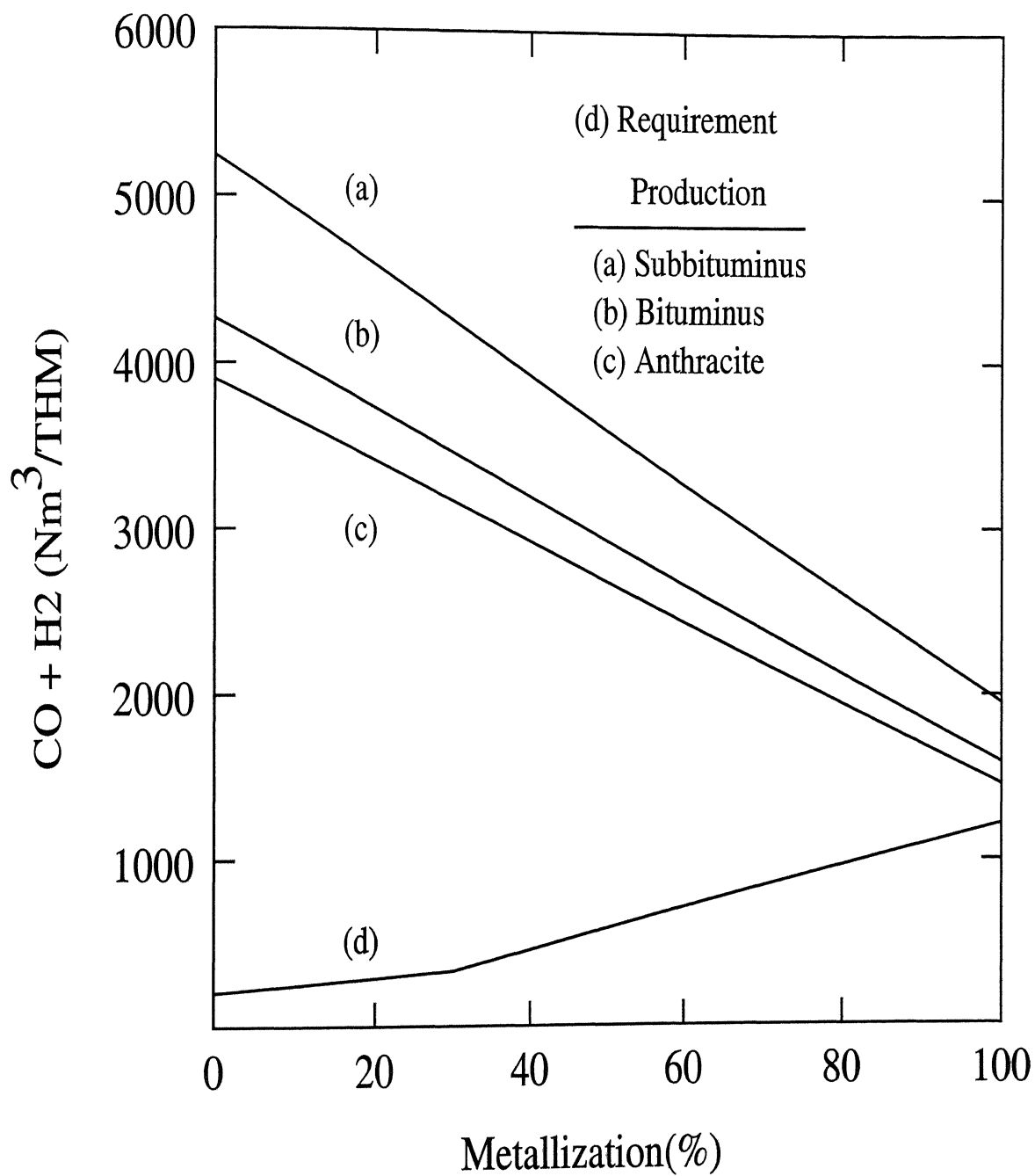


Figure 4.6: $(\text{CO} + \text{H}_2)$ production and requirement vs degree of metallization for various types of coal

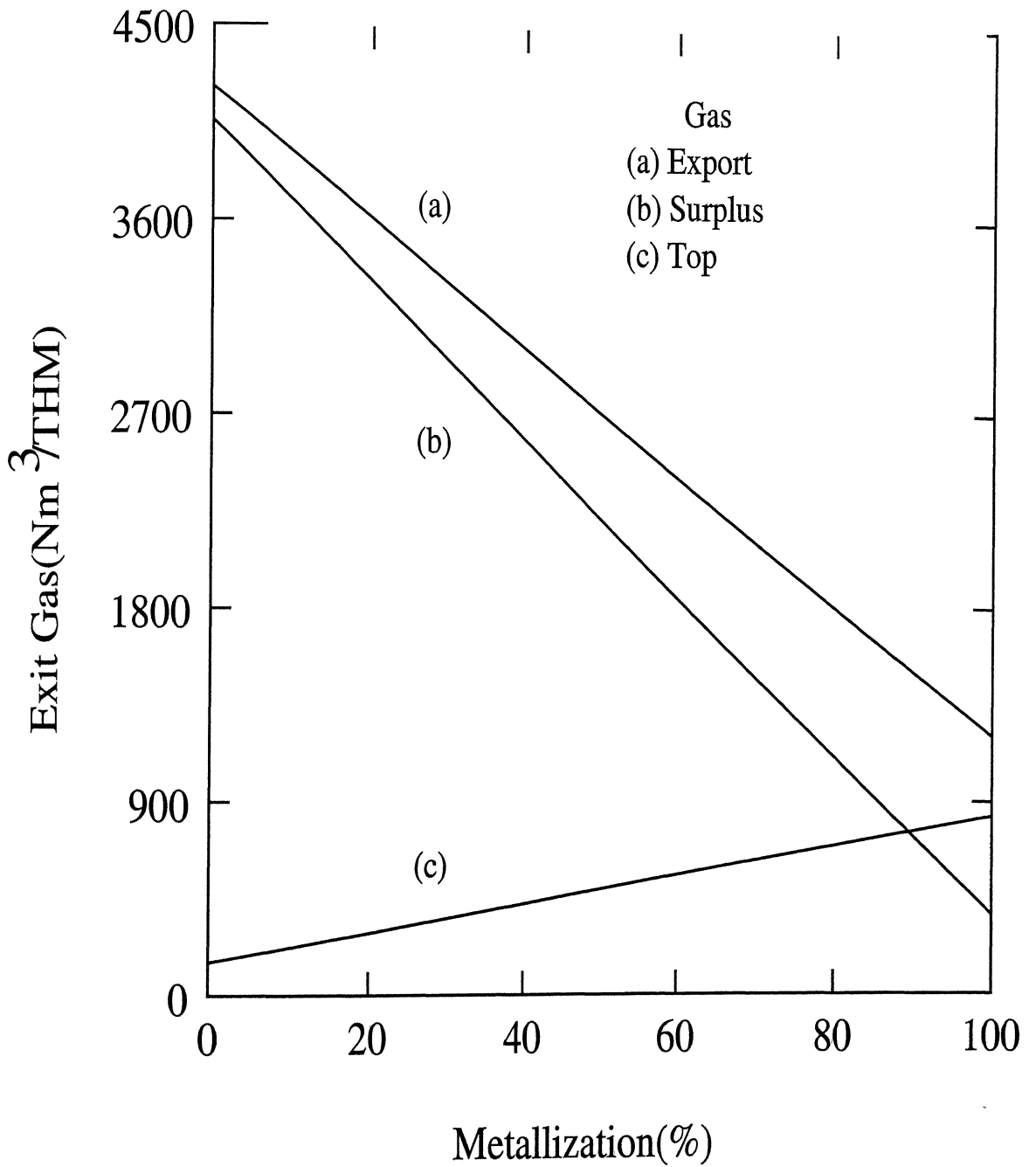


Figure 4.7: Exit gas vs degree of metallization

metallization increases from 0% to 100% shown in figure 4.7.

export gas is total of surplus gas and top gas. As surplus gas decreases faster than top gas increases with degree of metallization. So, ultimately export gas decreases with degree of metallization increases shown in figure 4.7. Export gas varies from $4212.2 \text{ Nm}^3/\text{thm}$ to $1569 \text{ Nm}^3/\text{thm}$ when metallization increases from 0% to 100%. Export gas($\text{CO} + \text{H}_2$) at different degrees of metallization is given in table 4.13.

4.2 Effect of Post Combustion Ratio

The gas produced during the combustion of coal in the melter-gasifier are CO , H_2 and N_2 . The post combustion of CO and H_2 is a very useful method for reducing coal consumption. Post combustion of CO and H_2 produces too much heat in the melter-gasifier and hence reduces coal consumption. For post combustion, necessary amount of oxygen or, air or, preheated air or, O_2 -enriched air are sent from the top of the melter-gasifier. Effect of post combustion ratio on degree of dissociation, gas temperature, coal rate, O_2 requirement, flux rate, slag rate and export gas are discussed in this section.

4.2.1 Dissociation and Gas Temperature

Post combustion of CO and H_2 produces too much heat in the melter-gasifier and increases the temperature of the gas. For this reason, dissociation of gas component occurs. In the range of post combustion ratio are considered(0%-50%) , only CO_2 dissociation occurs and H_2O dissociation does not occurs. In figure 4.8 degree of dissociation of CO_2 $\left(\left(\frac{\text{Moles of } \text{CO}_2 \text{ dissociated}}{\text{Moles of } \text{CO}_2 \text{ before dissociation}} \right) \times 100 \right)$ has been plotted against post combustion ratio for different post combustion medium. (see table 4.22 for various values of dissociation). When 40% O_2 -enriched air or, air at 298°K , 700°K and 1000°K are sent in the melter-gasifier for post combustion, then there are no dissociation occurs. When oxygen is used for post combustion, then temperature goes higher than when preheated air is sent.

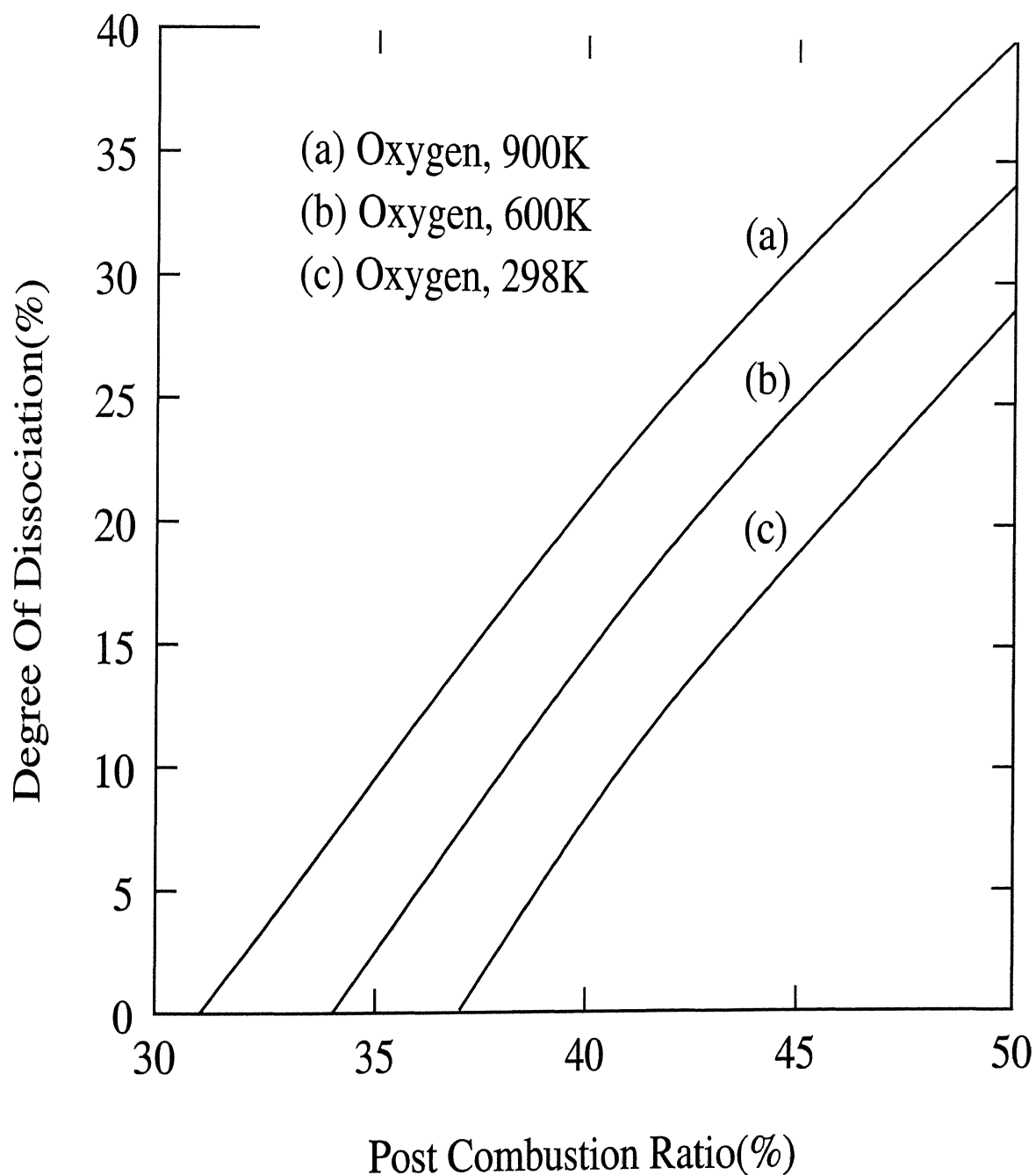


Figure 4.8: Degree of dissociation vs post combustion ratio for different post combustion medium

So, here dissociation starts at 37% post combustion ratio and dissociation increases as post combustion ratio increases. When oxygen is preheated to $600^{\circ}K$ or, $900^{\circ}K$, then dissociation starts at lower post combustion ratio. When preheated oxygen at $600^{\circ}K$ is sent for post combustion, then dissociation starts at 34% post combustion ratio and as the preheated temperature increases to $900^{\circ}K$, then dissociation starts at 31% post combustion ratio.

In figure 4.9 gas temperature has been plotted against post combustion ratio for different post combustion medium. In this figure we see that the temperature of the gas increases from $1273^{\circ}K$ to $3691^{\circ}K$ (see table 4.21) as we increase the post combustion ratio from 0% to 50% when oxygen is used for post combustion. From this figure we see that, temperature of gas is more for oxygen than temperature when even preheated air is used. This is because of increasing amount of nitrogen for preheated air. The star in the figure indicates the onset of dissociation of CO_2 starts. The temperature was supposed to decrease at a faster rate than shown in figure. But this is increasing only because of increasing H_2 consumption for post combustion after dissociation. In figure 4.10 it is shown when oxygen is used for post combustion. At 37% post combustion ratio CO_2 dissociation starts. For this reason, after 37% post combustion ratio, CO consumption decreases too much and comes near about same as at 50% post combustion ratio. But H_2 consumption increases more higher rate because at the time of CO_2 dissociation H_2 reacts with oxygen which is produced by dissociation. This is the reason why temperature of gas increases gradually after dissociation also.

4.2.2 $(CO + H_2)$ Production and Requirement

If the post combustion ratio is very high, then the gas will not have sufficient reduction potential to obtain the required degree of metallization. So, $(CO + H_2)$ production in the melter-gasifier and $(CO + H_2)$ requirement in the reduction shaft furnace gives important information. CO_2 and H_2O are removed from the gas goes to reduction shaft furnace for reduction. In figure 4.11 $(CO + H_2)$ production in the melter-gasifier and its requirement in the reduction shaft furnace have been plotted against degree of metallization for various

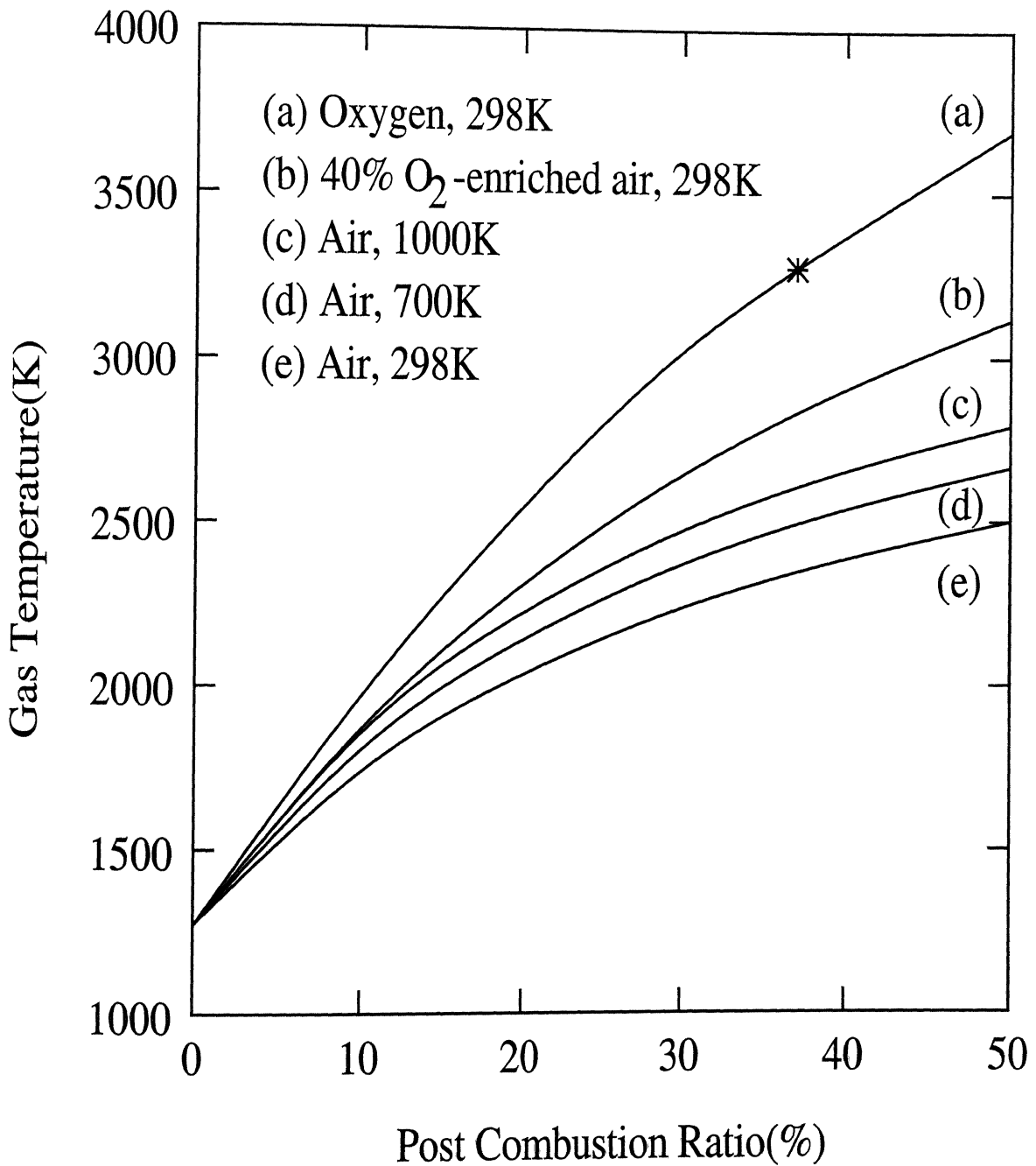


Figure 4.9: Gas temperature vs post combustion ratio for different post combustion medium

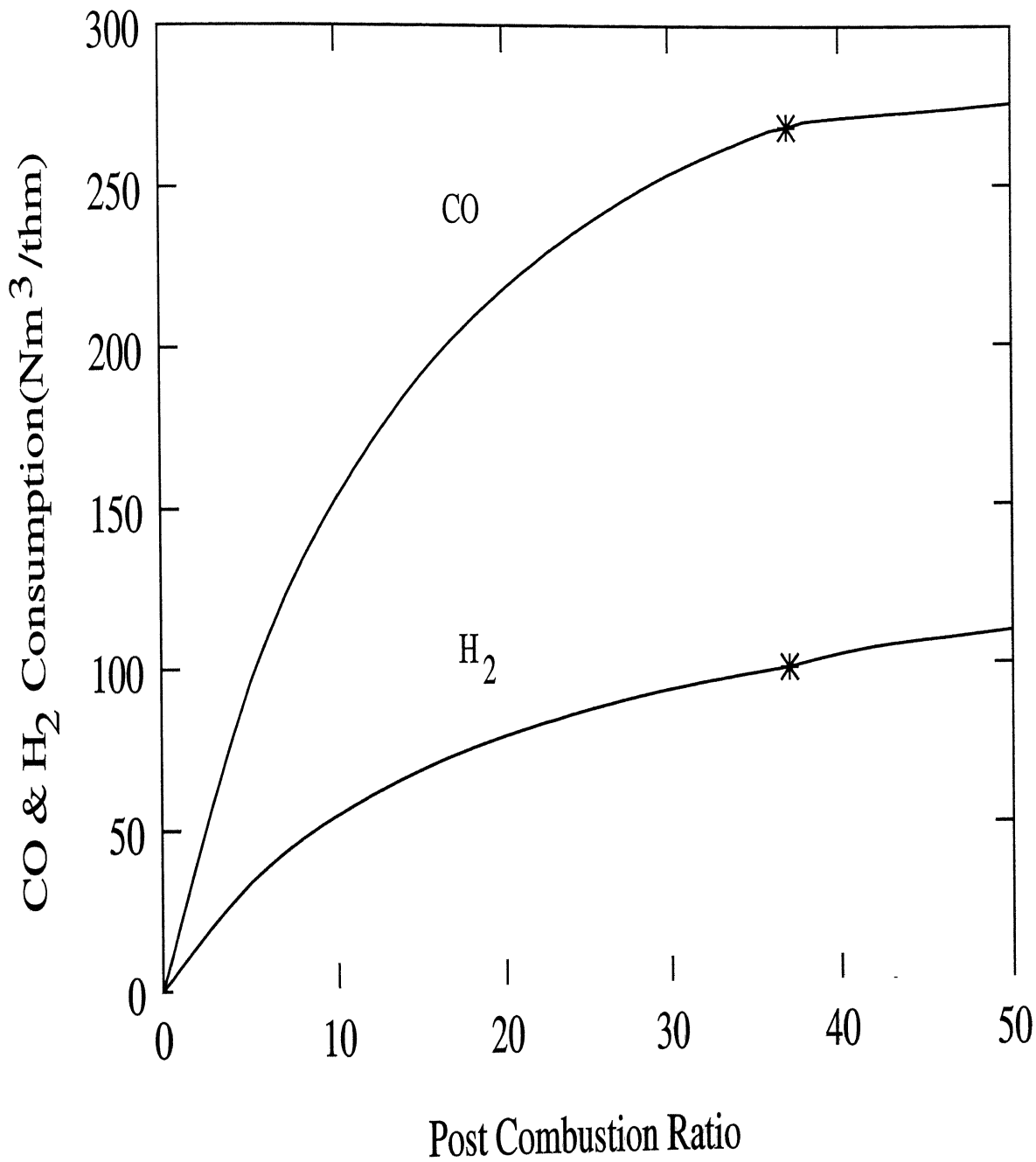


Figure 4.10: CO and *H*₂ consumption vs post combustion ratio

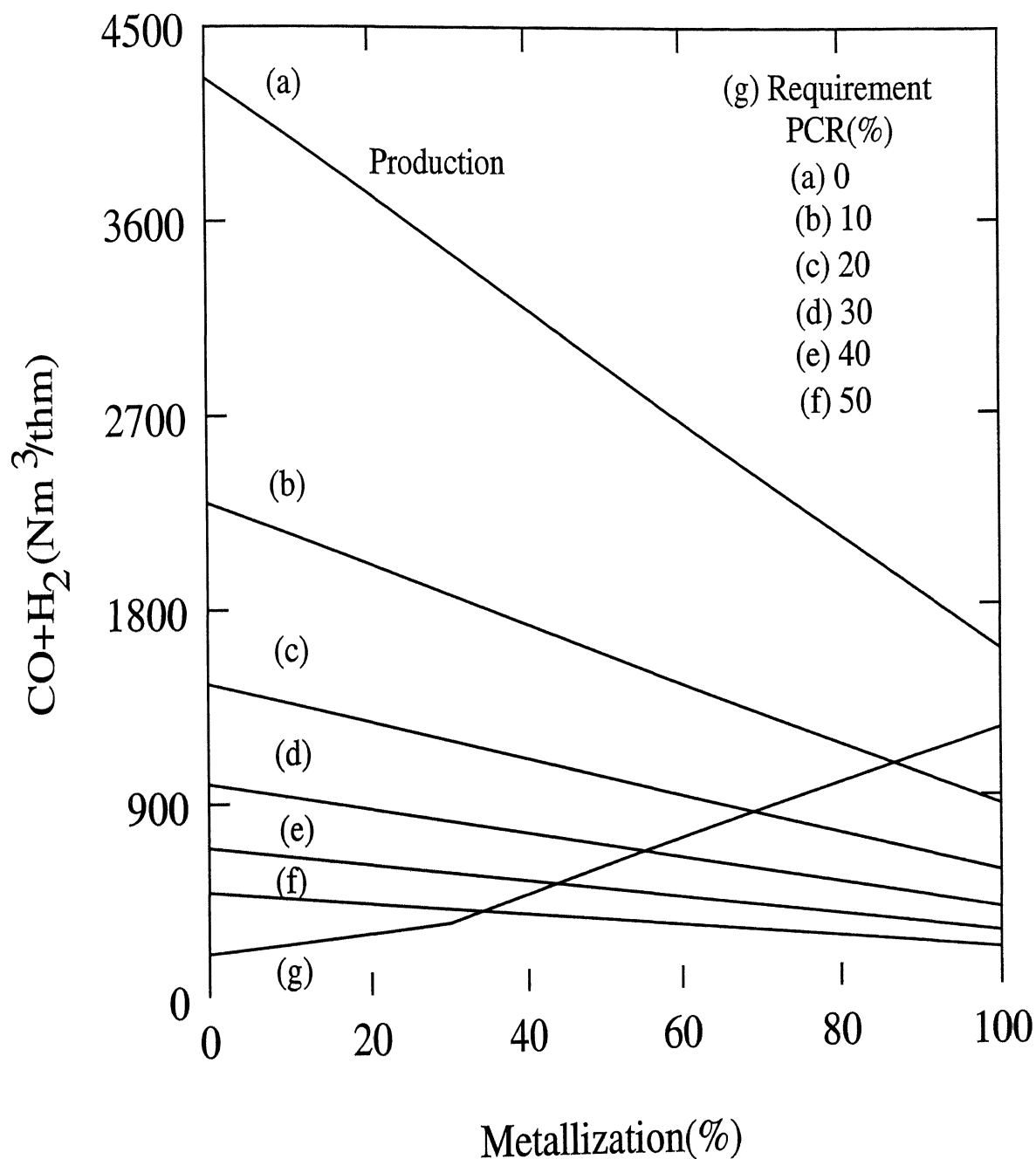


Figure 4.11: $(\text{CO} + \text{H}_2)$ production and requirement vs degree of metallization at various post combustion ratio

post combustion ratio. The figure shows that at higher post combustion ratio ($\text{CO} + \text{H}_2$) production in the melter-gasifier is less than the requirement of ($\text{CO} + \text{H}_2$) in the reduction shaft furnace as ($\text{CO} + \text{H}_2$) production decreases with post combustion ratio because of CO and H_2 consumption and conversion to CO_2 and H_2O respectively for post combustion. That means at higher post combustion ratio, high degree of metallization can not be achieved because the gas produced in the melter-gasifier will not have enough reduction potential to obtain that amount of metallization desired. For example, at 30% post combustion ratio, ($\text{CO} + \text{H}_2$) production and ($\text{CO} + \text{H}_2$) requirement curves cut before 60% metallization. So, at 30% post combustion ratio, 60% metallization can not be achieved. Optimum metallizations for different post combustion ratio are given in table 4.23.

Offcourse this is possible to reform the offgas, so that it could carry the reduction of ore according to the degree of metallization desired[15]. However, for reforming additional coal is required and produces more gas. Therefore the process being considered that the post combustion ratio and degree of metallization must be chosen so that the reforming is negligible.

4.2.3 Coal Rate

Coal consumption rate has been plotted against post combustion ratio for different degree of metallization in figure 4.12. Here, oxygen is used for post combustion, bituminous coal consumption and 100% heat transfer efficiency are considered. In this figure we see that with increasing post combustion ratio, coal consumption decreases for more heat production at higher post combustion ratio. At 0% metallization and 0% post combustion ratio coal consumption is 1977.4 Kg/thm and it decreases to 492.3 Kg/thm when post combustion ratio increases to 50%. At 30% metallization it is possible to achieve 50% post combustion ratio(see figure 4.11) and coal consumption decreases to 404.2 Kg/thm at 30% metallization and 50% post combustion ratio.

In case of 90% metallization with no post combustion, the coal consumption is consid-

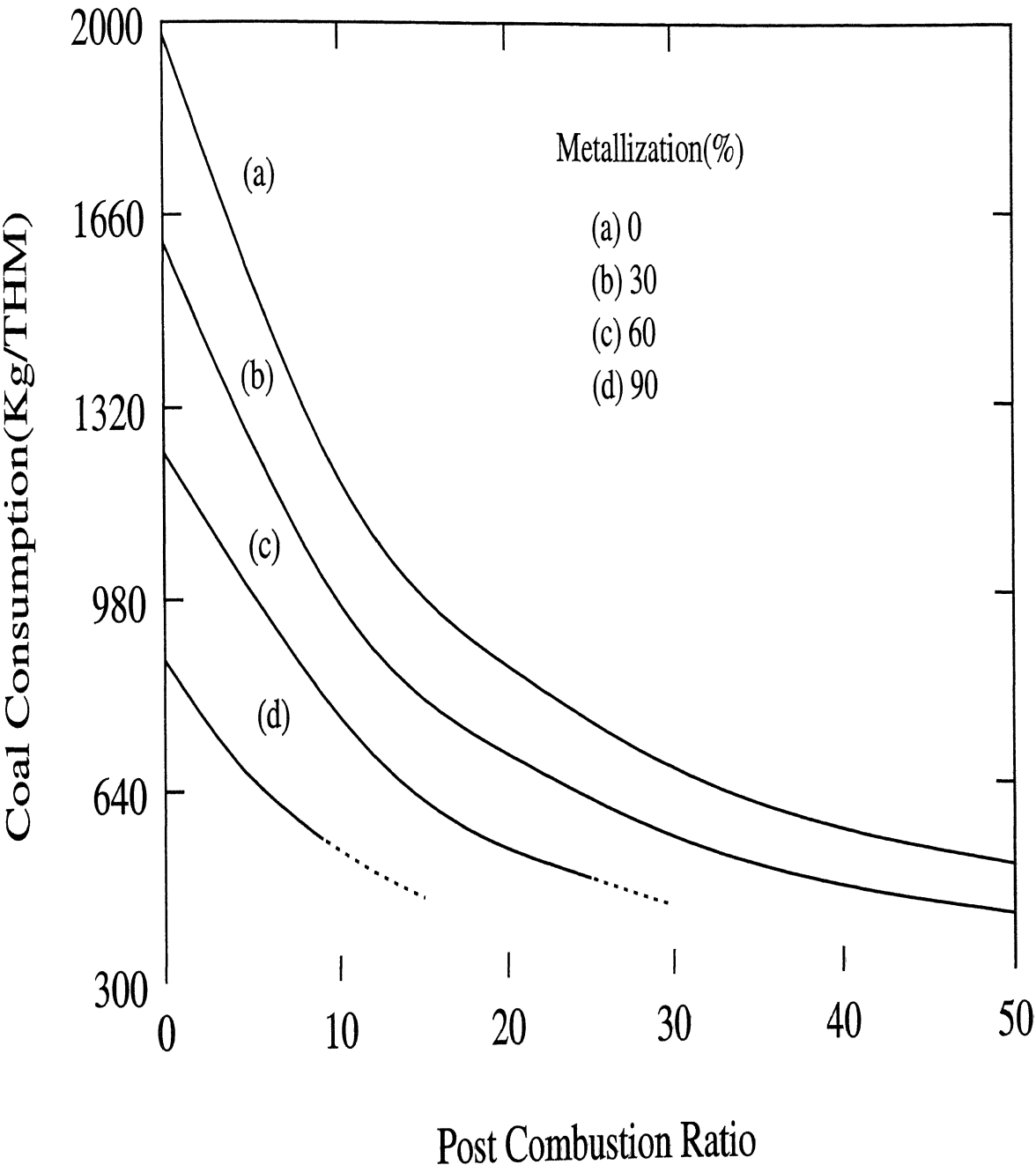


Figure 4.12: Coal consumption rate vs post combustion ratio for different degrees of metallization

erable greater than coal consumption for 30% metallization and 50% post combustion ratio. In this case, the melter-gasifier is being run under highly reducing conditions. With as little as at 10% post combustion ratio, the coal consumption decreases from 873.3 Kg/thm at 0% post combustion to 531.5 Kg/thm. However, the offgas from the melter-gasifier will not give 90% metallization without reforming as is called for because offgas is not suitable for that amount of reduction of iron ore. Dotted part of this curve is where offgas is not suitable for achieving the reduction without reforming or, scrubbing. For that the solid part of the curve for 90% metallization is not upto 10% post combustion ratio.

Similarly, in case of 60% metallization and 30% post combustion ratio the coal consumption is 426.4 Kg/thm. In this case also we can not get coal consumption upto that without reforming because at 60% metallization the offgas is not suitable beyond 25% post combustion ratio. For this, solid part of this curve is upto 25% post combustion ratio and dotted part is where the offgas is not suitable for reduction. The coal rate for different post combustion ratio is given in table 4.2.

Figure 4.13 shows coal saving $\left(\left(\frac{C_0 - C_{per}}{C_0} \right) \times 100 \right)$ at different heat transfer efficiency. As the post combustion increases, coal consumption decreases shown in figure 4.12. So, coal saving increases with post combustion ratio. When there is 100% heat transfer in the bath, then coal consumption at 0% post combustion ratio and 30% metallization coal consumption is 873.3 Kg/thm(C_0) and it decreases to 531.5 Kg/thm(C_{10}) at 10% post combustion ratio. So, at 10% post combustion ratio coal saving is 39.6% and it increases to 74.9% when post combustion ratio increases to 50%. Similarly for other heat transfer efficiency coal saving increases with post combustion ratio shown in this figure.

4.2.4 Oxygen Requirement

We know that oxygen requirement for coal is directly proportional to coal consumption. As the coal consumption decreases with post combustion ratio, oxygen requirement for coal also decreases with post combustion ratio. This is shown in figure 4.14. In this figure, oxygen

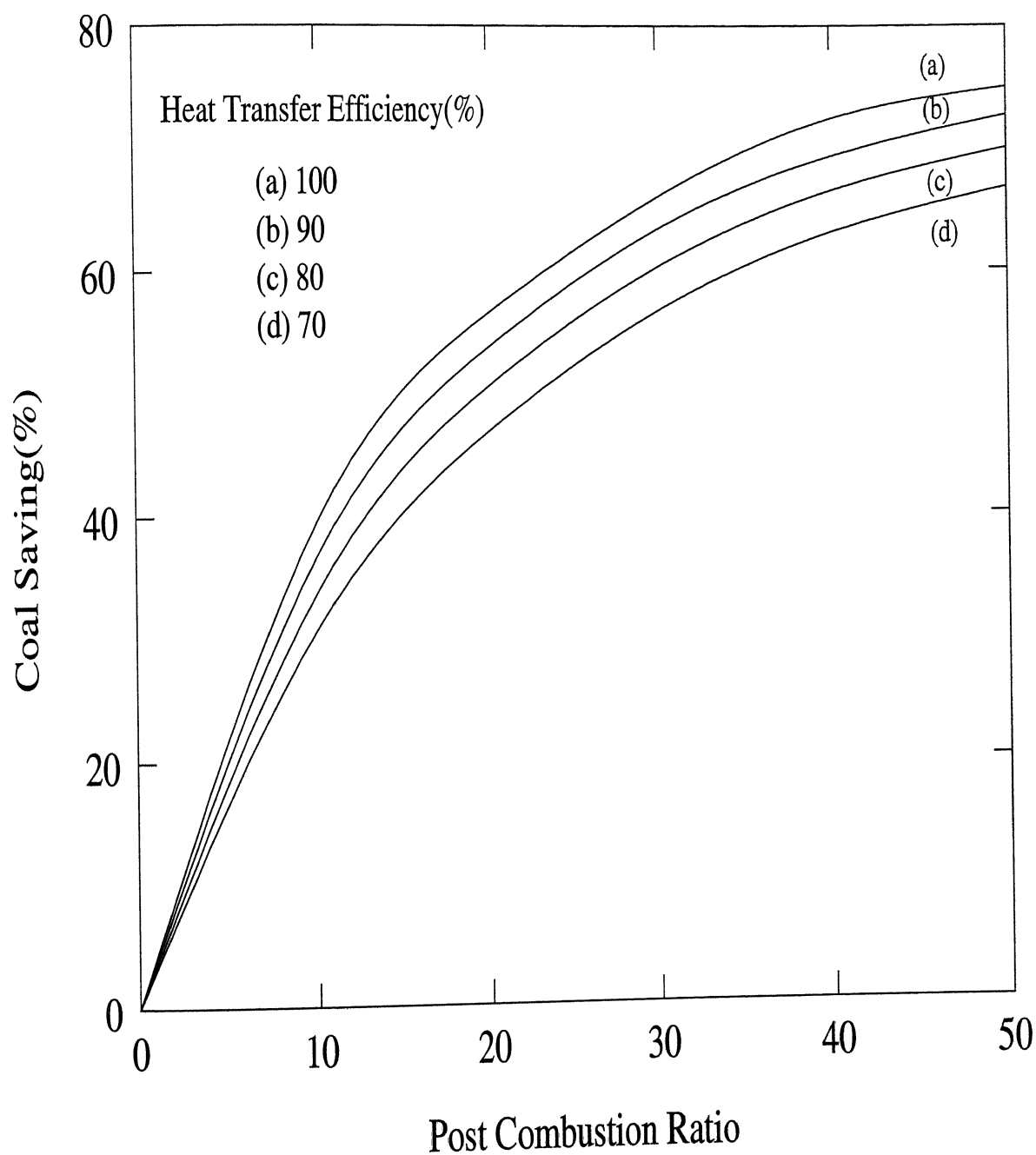


Figure 4.13: Coal saving vs post combustion ratio at different different heat transfer efficiency

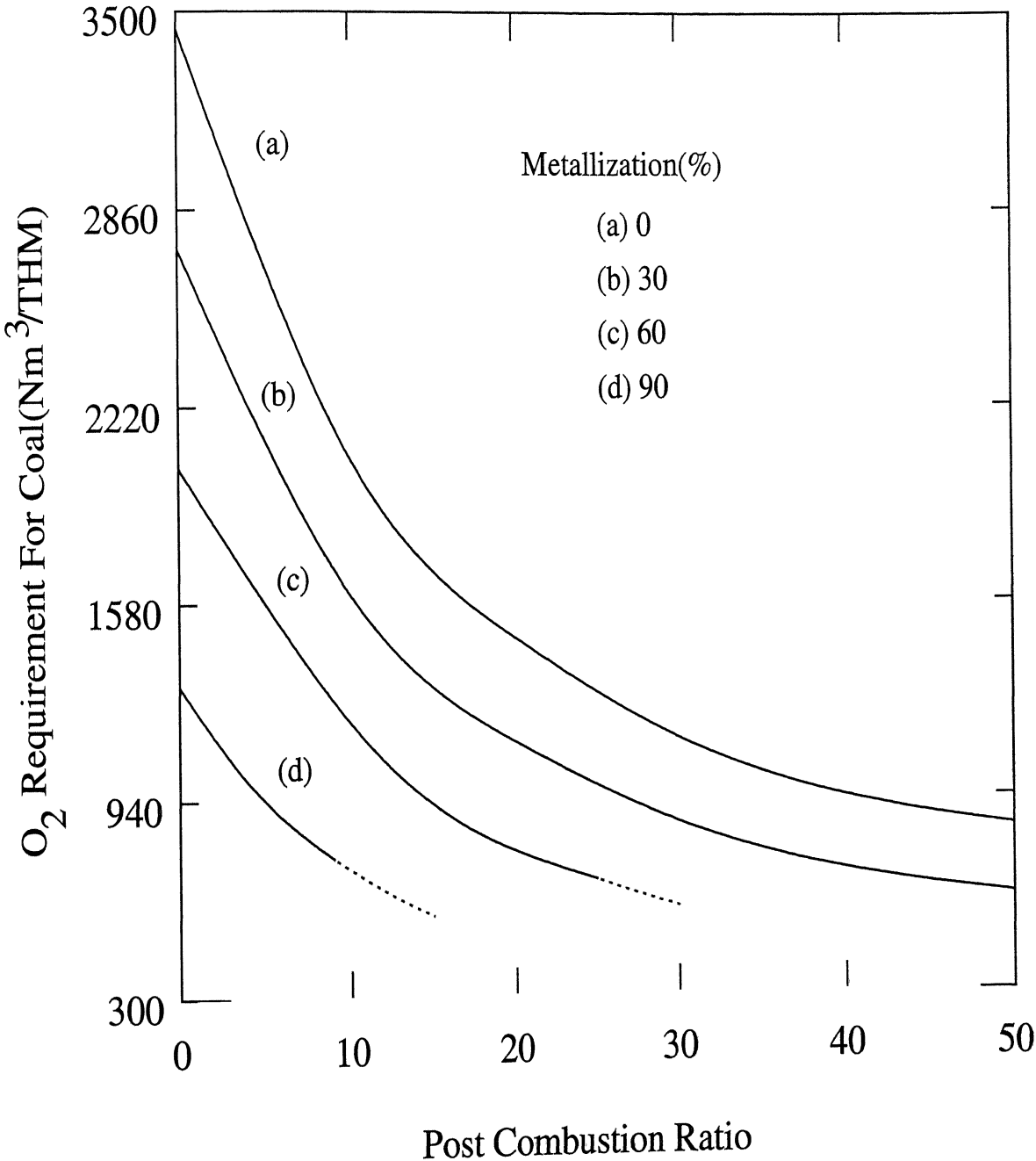


Figure 4.14: Oxygen required for coal vs post combustion ratio at degrees of metallization

requirement for coal has been plotted against post combustion ratio for different degree of metallization. At 0% metallization we can achieve upto 50% post combustion. oxygen requirement for coal decreases from $3444.3 \text{ Nm}^3/\text{thm}$ to $841.5 \text{ Nm}^3/\text{thm}$ (see table 4.5, 4.6, 4.7 for different oxygen requirement value for different medium) as we increase post combustion ratio from 0% to 50%. Similarly for 30% metallization to achieve 50% post combustion ratio no reforming required and oxygen required for coal varies from $2731.8 \text{ Nm}^3/\text{thm}$ to $617.4 \text{ Nm}^3/\text{thm}$ when we increase post combustion ratio from 0% to 50%. As we know for 60% metallization the offgas is suitable for reduction of ore upto 25% post combustion ratio. For this, the solid part of this curve is extended upto 25% post combustion ratio and the dotted curve is where offgas is not suitable for reduction without reforming described earlier also in coal rate subsection. Similar trend is seen for 90% post combustion ratio. This solid part of this curve is extended upto 9% post combustion ratio for the same reason.

In figure 4.15 oxygen required for post combustion has been plotted against post combustion ratio for different degrees of metallization. We see from this figure that with post combustion ratio, oxygen requirement for post combustion increases with post combustion ratio because more CO and H_2 are consumed for post combustion with post combustion ratio increases. At 0% metallization oxygen requirement for post combustion increases from 0 to $514.1 \text{ Nm}^3/\text{thm}$ (see table 4.5) as we increase post combustion ratio from 0% to 50%. similar trend are seen for other degree of metallization. For 90% metallization the solid part of the curve is extended upto 9% metallization (explained earlier).

Oxygen requirement for coal is decreasing at a faster rate than oxygen required for post combustion. As we increase post combustion ratio upto 10%, oxygen required for coal decreases from $3444.9 \text{ Nm}^3/\text{thm}$ to $2055.4 \text{ Nm}^3/\text{thm}$ i.e. $989.5 \text{ Nm}^3/\text{thm}$ oxygen requirement decreases. Oxygen required for post combustion increases from 0 to $255 \text{ Nm}^3/\text{thm}$ i.e. $255 \text{ Nm}^3/\text{thm}$ oxygen required increases. So, ultimately total oxygen requirement decreases with post combustion ratio.

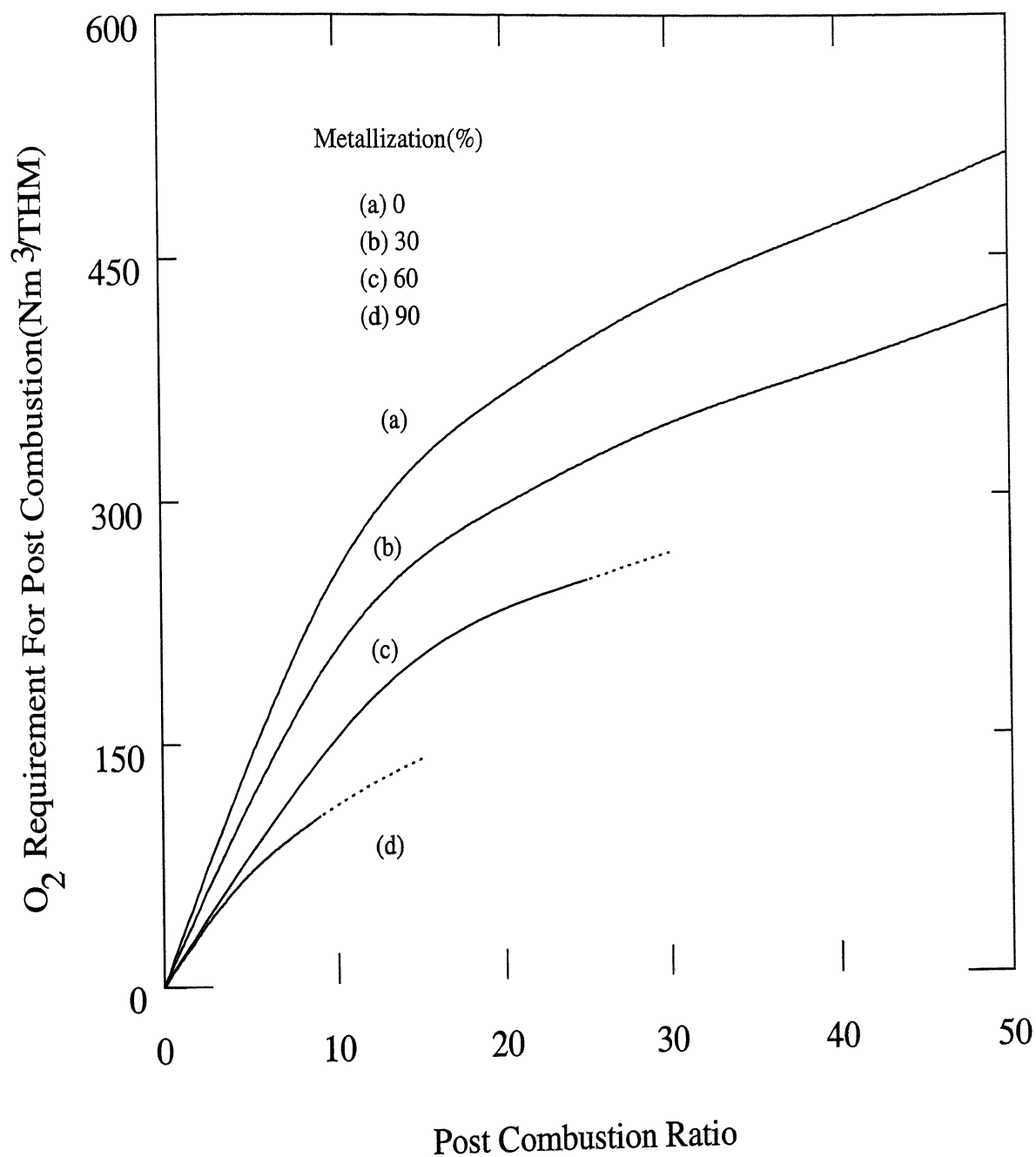


Figure 4.15: Oxygen required for post combustion vs post combustion ratio at different degrees of metallization

4.2.5 Flux and Slag Rate

Flux rate at different post combustion ratio is shown in figure 4.16. In this figure we see that as the post combustion ratio increases, flux rate decreases. This is because of maintaining basicity 1.2. As the post combustion ratio increases, coal consumption decreases (described earlier). So, less amount of SiO_2 and Al_2O_3 comes from coal. As the basicity of the slag is maintained 1.2, less amount of CaO and MgO comes from flux and hence decreases flux rate. At a certain metallization (let's 30% metallization) flux rate is 260.3 Kg/thm at 0% post combustion ratio and it decreases to 192.7 Kg/thm when post combustion ratio increases to 50% post combustion ratio. Flux rate for different post combustion ratio and degree of metallization are given in table 4.8.

As, at higher post combustion ratio less SiO_2 and Al_2O_3 comes from coal and less CaO and MgO comes from flux. So, slag produced is low at higher post combustion ratio. Slag contains CaO, MgO, SiO_2 , Al_2O_3 , CaS and MnO. All these component of slag except MnO decreases with post combustion ratio increases for less coal and flux. Decrease of slag rate with post combustion at different degrees of metallization is shown in figure 4.17. At 30% metallization and at 0% post combustion ratio slag rate is 284.1 Kg/thm and it decreases to 208.3 Kg/thm at 50% post combustion ratio (see table 4.9).

4.2.6 Export Gas

Export gas is used for many purposes like for power generation, production of DRI, heating purposes within an iron and steel complex, synthesis gas generation etc.

$$Export = Surplus + Top$$

As we know with post combustion ratio increases, $(CO + H_2)$ production decreases because CO_2 and H_2O produced which are removed by gas scrubber, but its requirement in the reduction shaft furnace does not vary with post combustion ratio because $(CO + H_2)$ sent in the reduction shaft furnace according to the degree of metallization desired. For this reason,

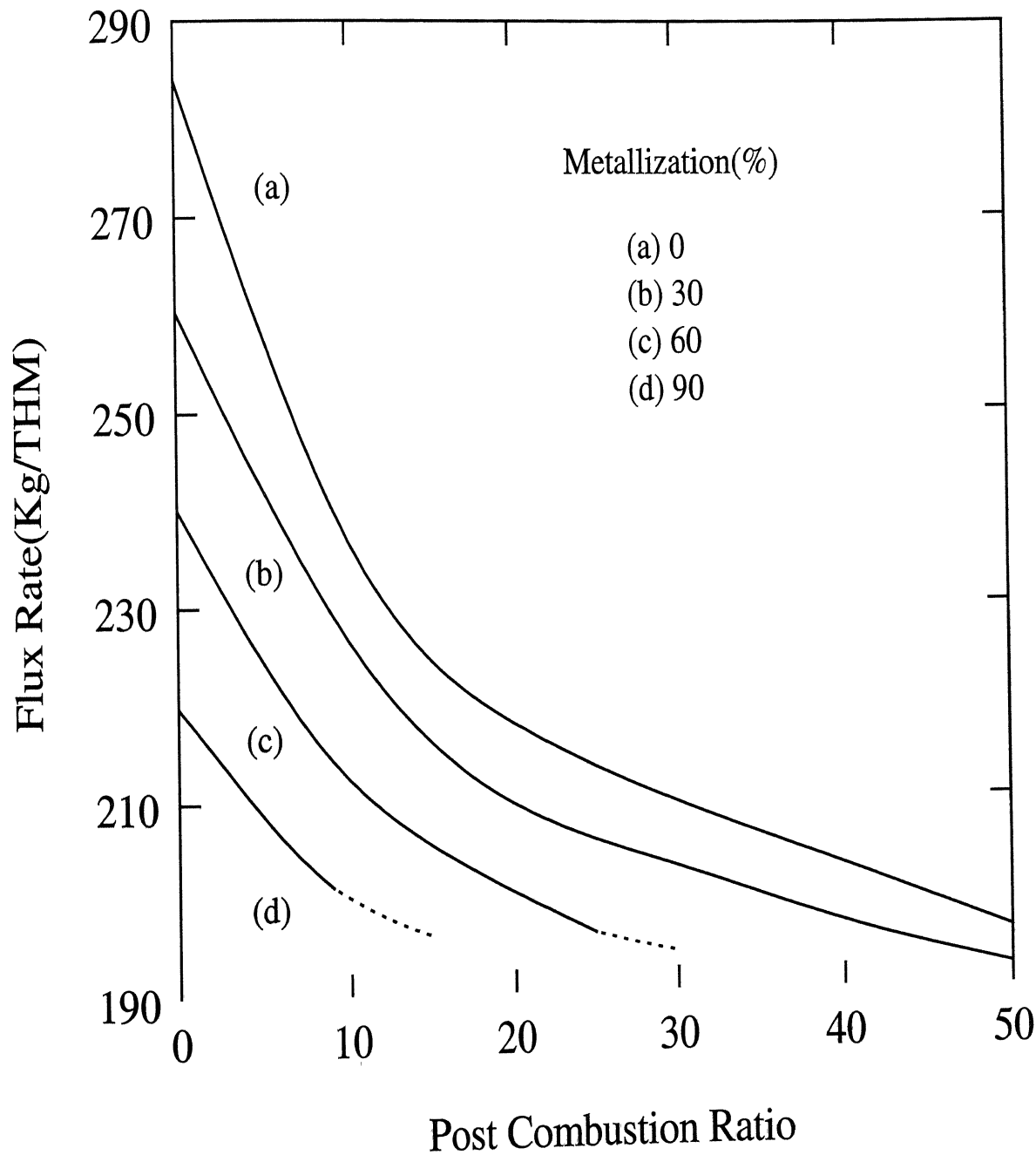


Figure 4.16: Flux rate vs post combustion ratio at different degrees of metallization

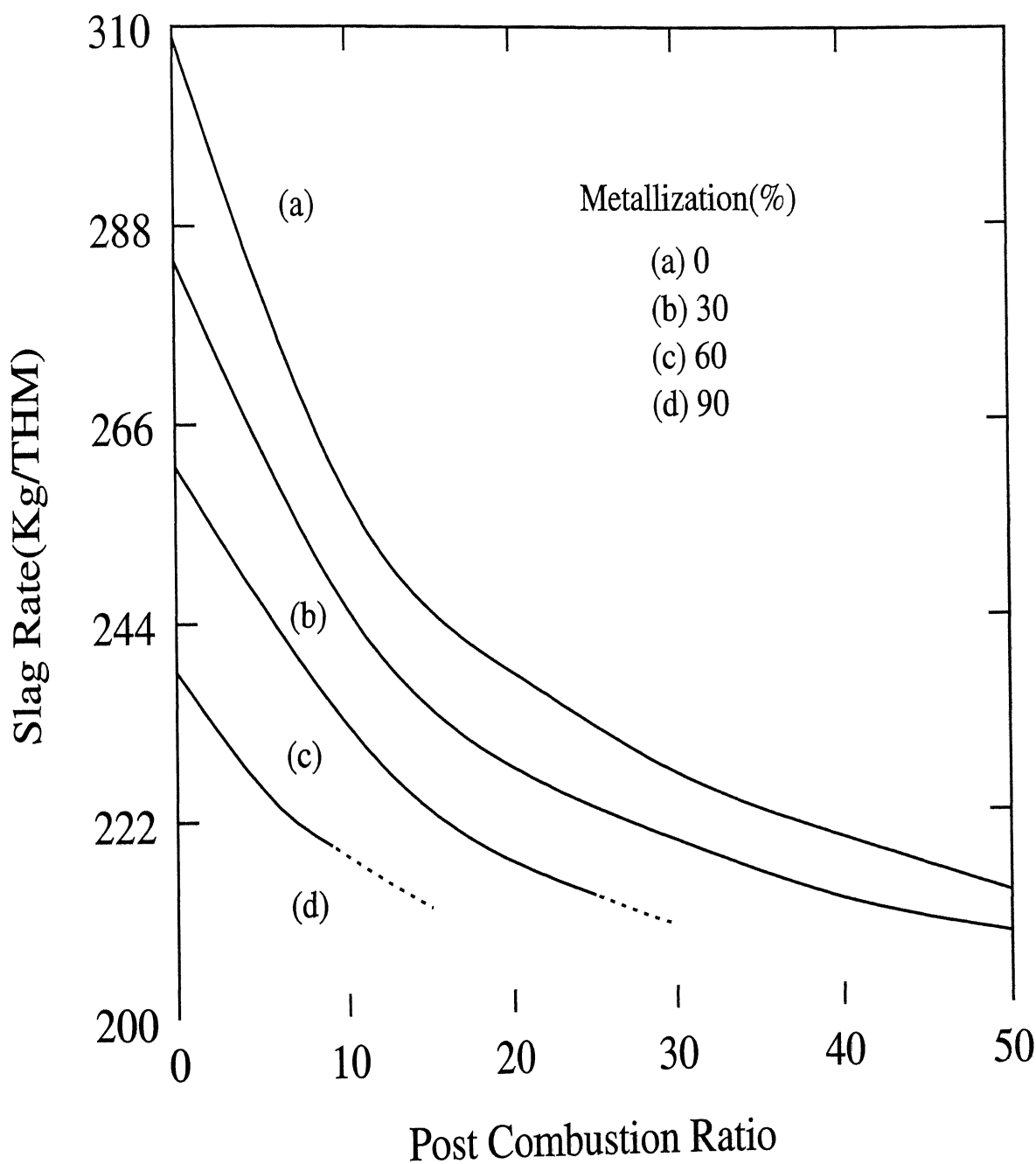


Figure 4.17: Slag rate vs post combustion ratio at different degrees of metallization

top gas amount does not vary with post combustion ratio is shown in figure 4.18. Top gas amount at various degrees of metallization are already shown and described in figure 4.7.

In figure 4.19 surplus gas has been plotted against post combustion ratio for different degrees of metallization.

$$\text{Surplus} = \text{Production} - \text{Requirement} - \text{Post combusted product}$$

As $(\text{CO} + \text{H}_2)$ produced by coal combustion post combusted according to post combustion ratio, so surplus gas is decreasing with post combustion ratio. At 60% metallization as $(\text{CO} + \text{H}_2)$ production after post combustion match the requirement in the reduction shaft furnace at 25% post combustion. That means at 25% post combustion ratio there are no gas comes out from the melter-gasifier. For this at this post combustion ratio surplus gas is zero. Similar trend is seen for 90% metallization also.

As export gas is total of surplus gas and top gas. So, export gas is decreasing with post combustion ratio shown in figure 4.20. In this figure we see that at 30% metallization, export gas decreases from $3319.9 \text{ Nm}^3/\text{thm}$ to $262 \text{ Nm}^3/\text{thm}$ as the post combustion ratio increases from 0% to 50%. We already know that at 60% metallization we cannot go above 25% post combustion ratio without reforming, this curve is upto 25% post combustion ratio. After this wustite-Fe reaction will not occurs. For 90% metallization we can go upto 9% post combustion ratio and export gas amount is $701.1 \text{ Nm}^3/\text{thm}$ at 9% post combustion ratio. Similarly for other degrees of metallization also shown in this figure. Surplus and export gas for different post combustion ratio and post combustion medium are given in table 4.10-4.15.

4.3 Post Combustion Medium

4.3.1 Gas Temperature and Volume

Post combustion medium is one important parameter to control gas volume in the melter-gasifier and hence temperature of the gas. A mixture of O_2 and air is used for post combustion

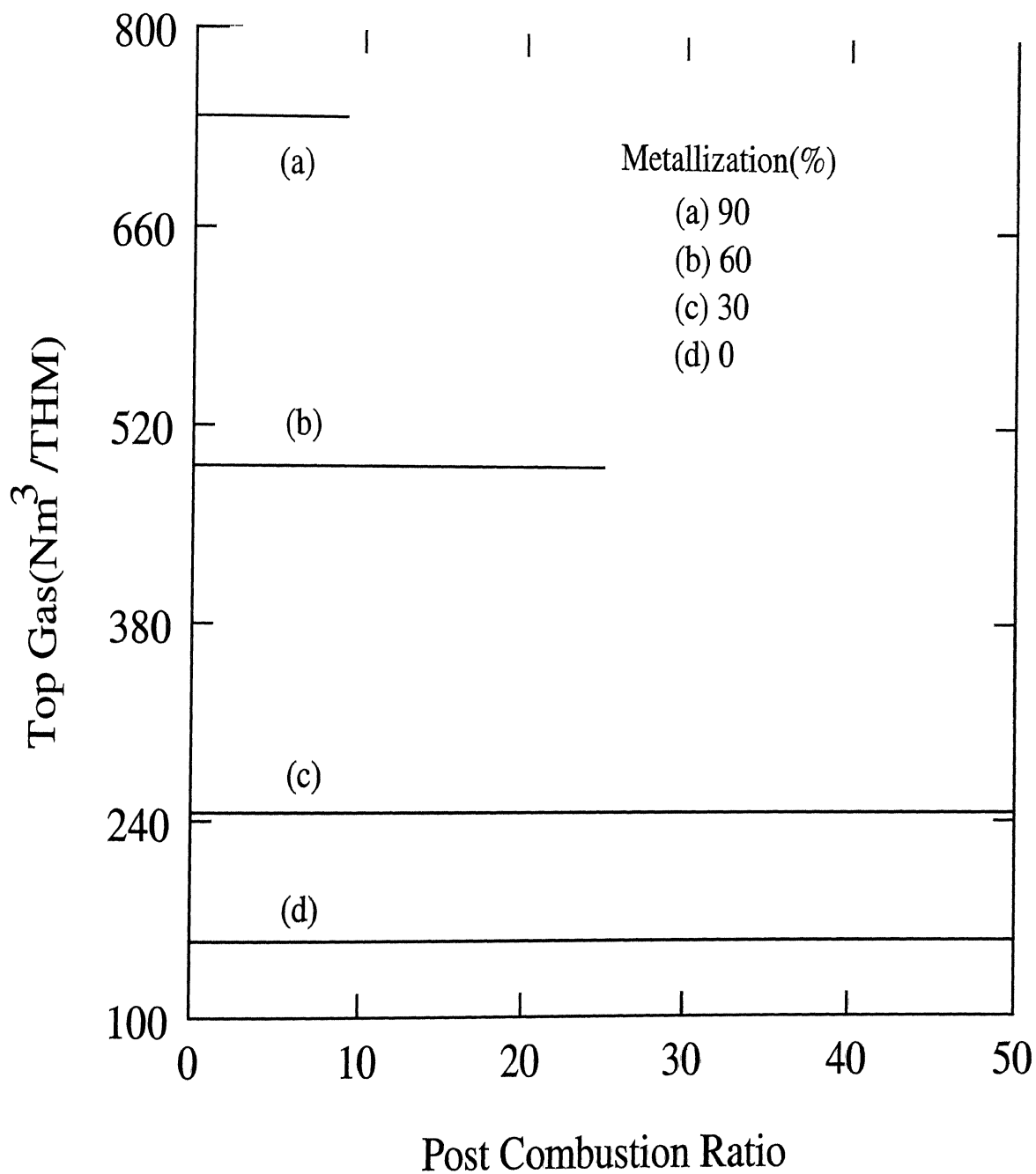


Figure 4.18: Top gas vs post combustion ratio at different degrees of metallization

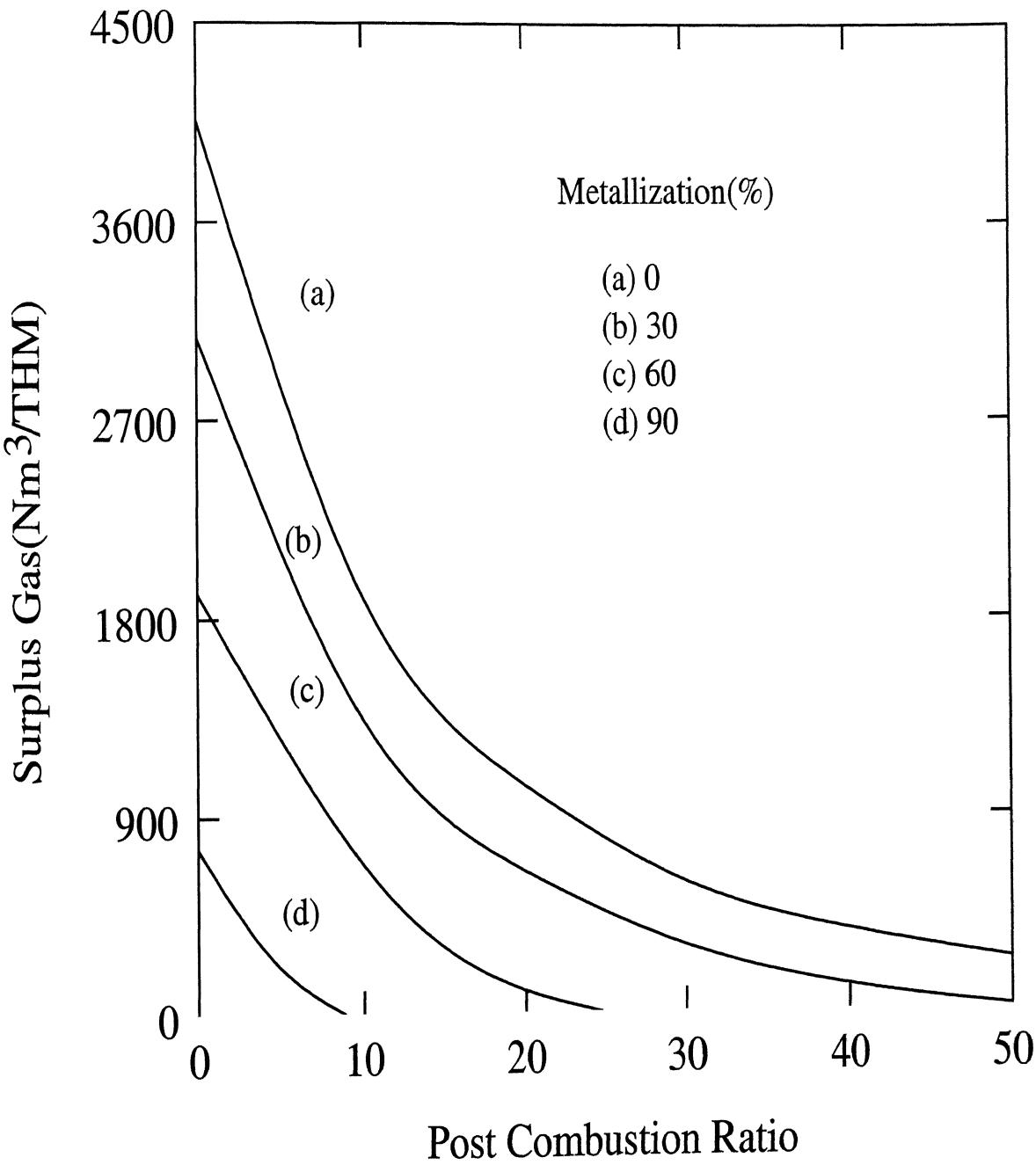


Figure 4.19: Surplus gas vs post combustion ratio at different degrees of metallization

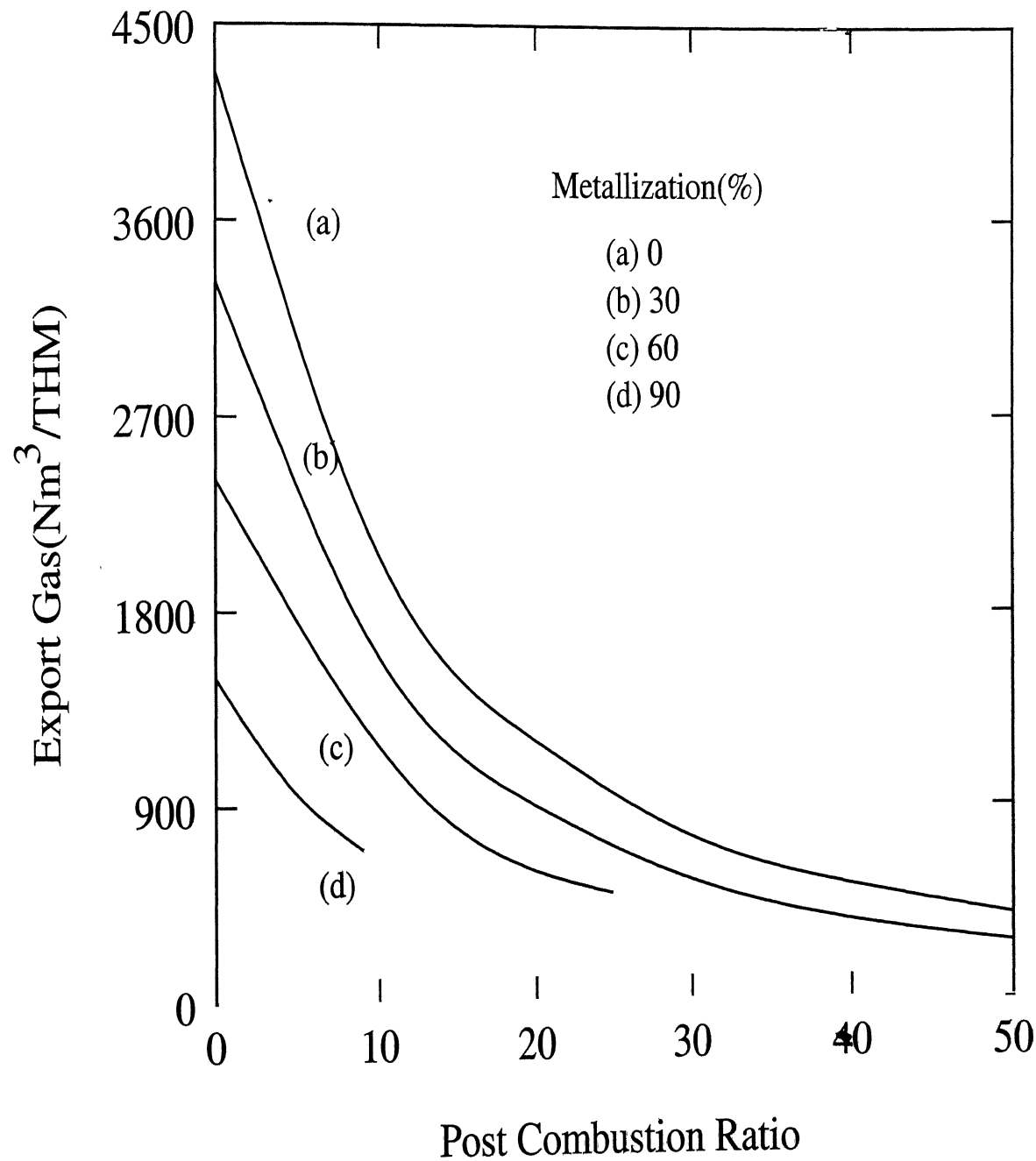


Figure 4.20: Export gas vs post combustion ratio at different degrees of metallization

to temper the temperature of the gas. In figure 4.21 medium required for post combustion has been plotted against post combustion ratio for different medium. In this figure we see that when oxygen is used for post combustion, then less amount of oxygen is used for post combustion. But, when 40% O_2 -enriched air is used, then nitrogen also comes in the melter-gasifier along with oxygen from this medium. For a certain percent post combustion, oxygen required for the combustion is near about same for all medium, but for nitrogen volume of the medium required is more when this O_2 -air mixture is used for post combustion. When air is used i.e. 21% oxygen and 79% nitrogen comes from this medium, so more volume is required for this. In this figure we see that for 10% post combustion ratio, oxygen required for post combustion is $207 \text{ Nm}^3/\text{thm}$, 40% O_2 -enriched air required $537.2 \text{ Nm}^3/\text{thm}$ and air required $1084 \text{ Nm}^3/\text{thm}$ (see table 4.16-4.18). Similarly for other post combustion ratio is shown in this figure. For this more volume of medium, gas volume in the melter-gasifier increases and temperature of the gas decreases.

This is shown in the figure 4.22. In this figure volume of gas produces ($CO + CO_2 + H_2 + H_2O + N_2$) and its temperature have been plotted against mole fraction of oxygen in the post combustion medium. This figure is drawn by considering 80% heat transfer in the bath and 50% post combustion ratio. In this figure it is clear that volume of the gas decreases with increasing fraction of oxygen in the post combustion medium. This is due to the increasing nitrogen content in the gas. As the fraction of oxygen increases, then less amount of nitrogen comes from post combustion medium. So, from the figure 4.22 we see that with increasing fraction of oxygen in post combustion medium increases, less gas produced and hence gas temperature increases because of absorbing heat for post combustion by less amount of gas. In this figure temperature of the gas is $1784^\circ K$, when 100% oxygen is used. This temperature can be lowered by using 40% O_2 -enriched air to $1590^\circ K$. If we use air i.e. .21 mole fraction of oxygen for post combustion, then the temperature decreases to $1547^\circ K$. The volume of the gas is $1446 \text{ Nm}^3/\text{thm}$ at 30% metallization, when 40% O_2 -enriched air is

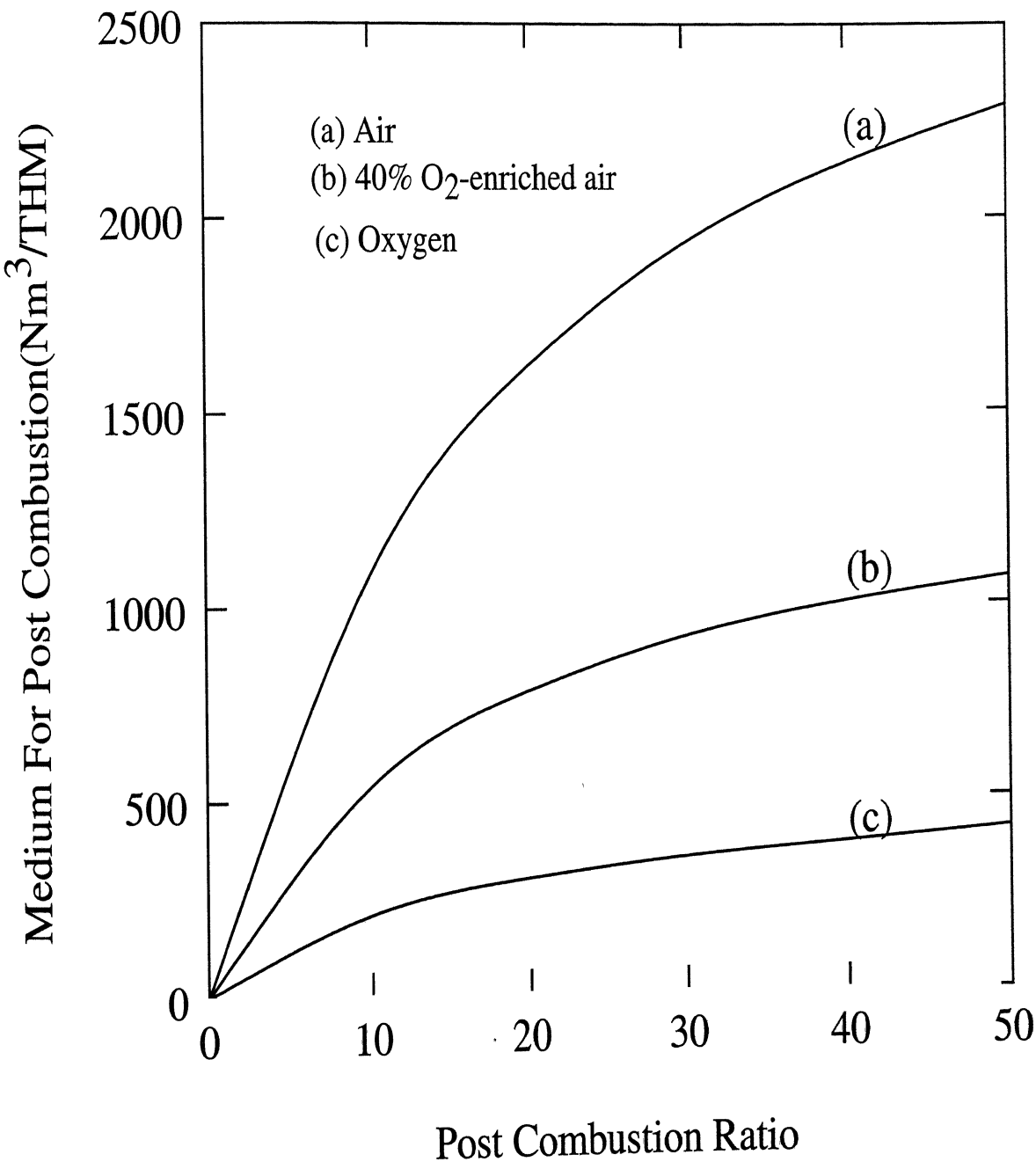


Figure 4.21: Medium required for post combustion post combustion ratio for different post combustion medium

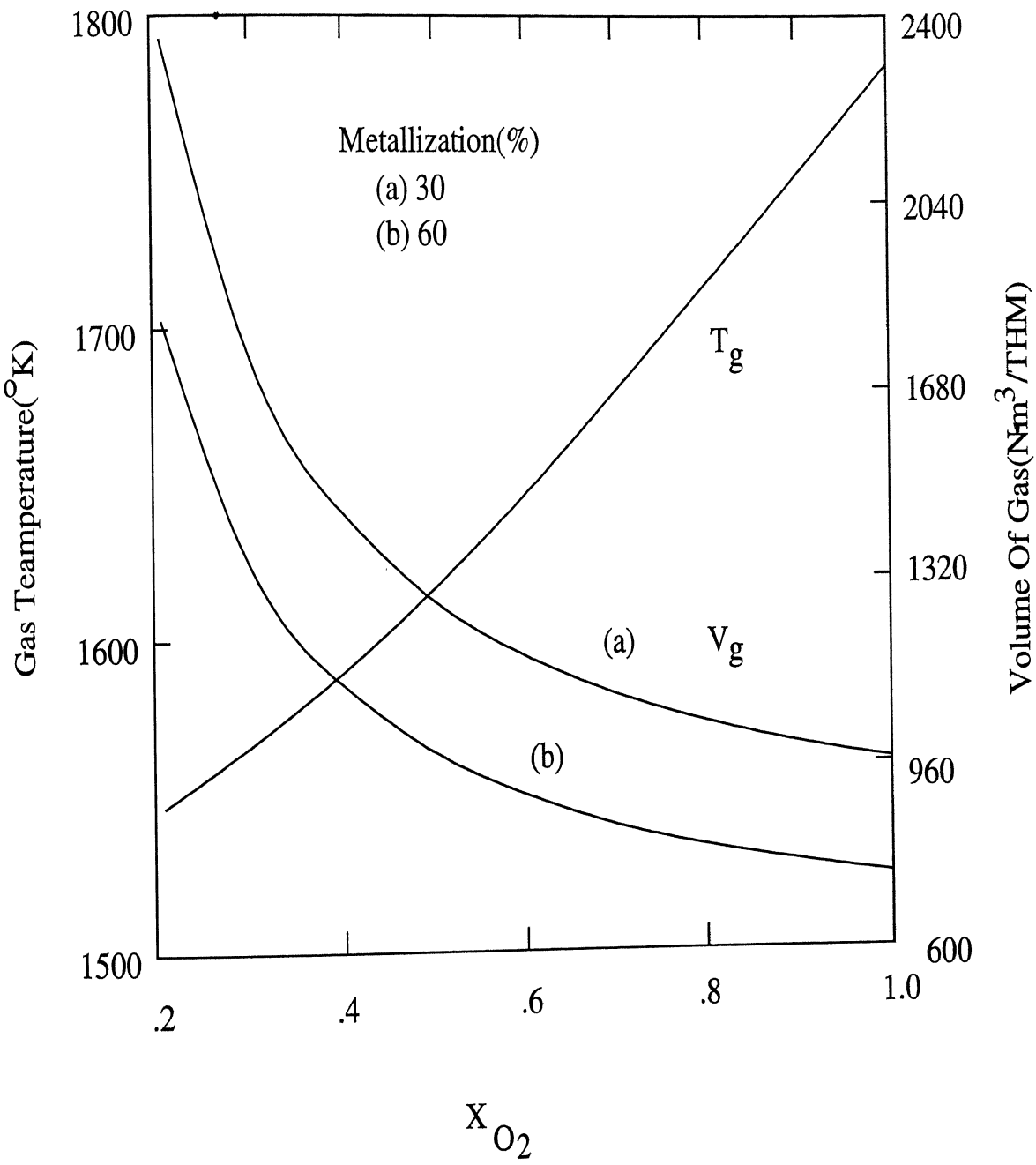


Figure 4.22: Volume of gas and its temperature vs mole fraction of oxygen in post combustion medium at different metallization

used for post combustion. Volume of gas for different post combustion medium is given in table 4.19. Volume of the gas increases to $2355.2 \text{ Nm}^3/\text{thm}$ at 30% metallization when air is used for post combustion. So, by controlling fraction of O_2 in post combustion medium, gas temperature can be controled.

4.3.2 Coal Rate

In figure 4.23 coal consumption rate has been plotted against post combustion ratio for different post combustion medium like O_2 , 40% O_2 -enriched air and air. This figure is drawn by considering 30% metallization, 100% heat transfer efficiency and bituminous coal composition. In this figure we see that when oxygen is used for post combustion, then coal consumption rate is lower than when 40% O_2 -enriched air is used for post combustion. When there is no post combustion, then coal consumption rate is same i.e. 1606.9 Kg/thm for all post combustion medium because no medium is sent for post combustion. This coal consumption is decreases to 404.2 Kg/thm when post combustion ratio increases to 50%.

When 40% O_2 -enriched air is used for post combustion, then 40% oxygen and 60% nitrogen comes in the melter-gasifier. As post combustion of CO and H_2 takes oxygen as for 100% oxygen, total volume of air increases has to be sent in the melter-gasifier. As amount of nitrogen comes more, it will take away more heat from melter-gasifier. For this reason, heat demand for melter-gasifier increases and hence coal consumption increases. At 50% post combustion ratio coal consumption increases to 415.5 Kg/thm for 40% O_2 -enriched air from 404.2 Kg/thm when oxygen is used for post combustion.

When air is sent for post combustion, then coal consumption is highest among these three mediums because more nitrogen comes from this medium. As only 21 volume percent oxygen comes from air, more volume of medium required for post combustion. Hence amount of nitrogen in the melter-gasifier is more than when 40% O_2 -enriched air is sent. Coal consumption increases to 453.9 Kg/thm at 50% post combustion ratio. When oxygen is sent for post combustion, then at 37% post combustion ratio CO_2 dissociation starts and

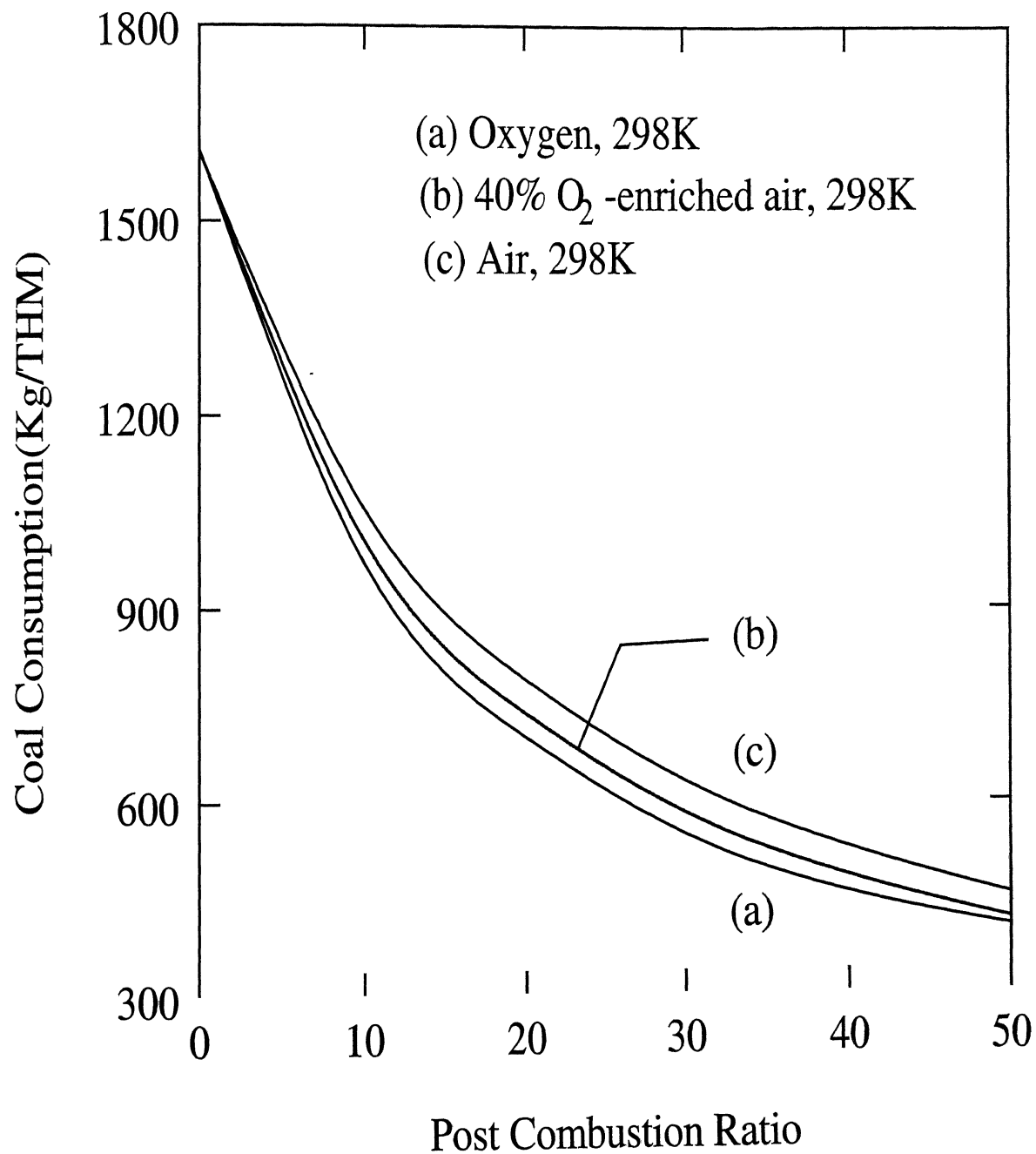


Figure 4.23: Coal consumption vs post combustion ratio for different post combustion medium

increases coal consumption. For this reason the difference between coal consumption for 40% O_2 -enriched air and for oxygen decreases as we increase post combustion ratio after 37% post combustion ratio. Coal consumption for various medium are given in table 4.2-4.4.

4.4 Effect of Heat Transfer Efficiency

Effect of heat transfer efficiency on gas temperature and coal consumption rate is discussed in this section :

4.4.1 Gas Temperature

Heat transfer affects gas temperature in the melter-gasifier. If heat transfer to the bath is less, then gas temperature goes too high and the top of the melter-gasifier may burn. So, heat transfer to the bath must be as high as possible. In figure 4.24 and 4.25, gas temperature has been plotted against heat transfer efficiency for different post combustion ratio when oxygen and $1000^\circ K$ preheated air are used for post combustion respectively. From figure 4.24 we see that when there is no heat transfer at 50% post combustion ratio, then temperature of the gas is too high i.e. $3691^\circ K$. The temperatures are $3373^\circ K$, $3018^\circ K$ and $2541^\circ K$ for 40%, 30% and 20% post combustion ratio respectively. This temperature decreases when heat transfer to the bath increases. Now, if the allowable temperature of the gas in the melter-gasifier is $2000^\circ K$, then the heat transfer efficiency should be above 44%, 60%, 67% and 71% for 20%, 30%, 40% and 50% post combustion is done respectively.

If $1000^\circ K$ preheated air is used for post combustion, similarly heat transfer to the bath should be controled in such a way that temperature of the gas does not go beyond $2000^\circ K$ and temperature are shown in figure 4.25. Gas temperature for different post combustion medium at different heat transfer efficiency is given in table 4.21.

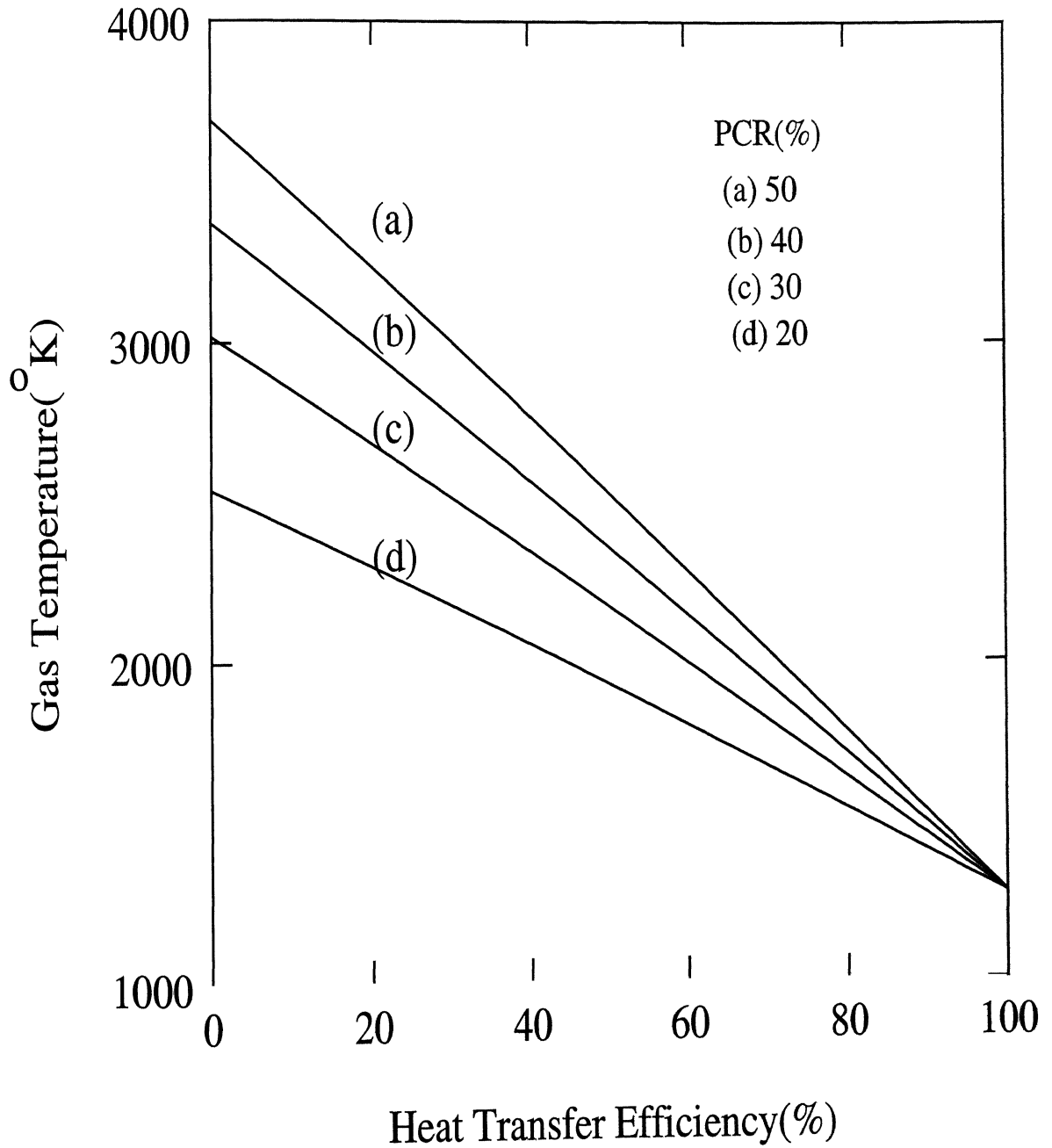


Figure 4.24: Gas temperature vs heat transfer efficiency at different post combustion ratio for oxygen as post combustion medium

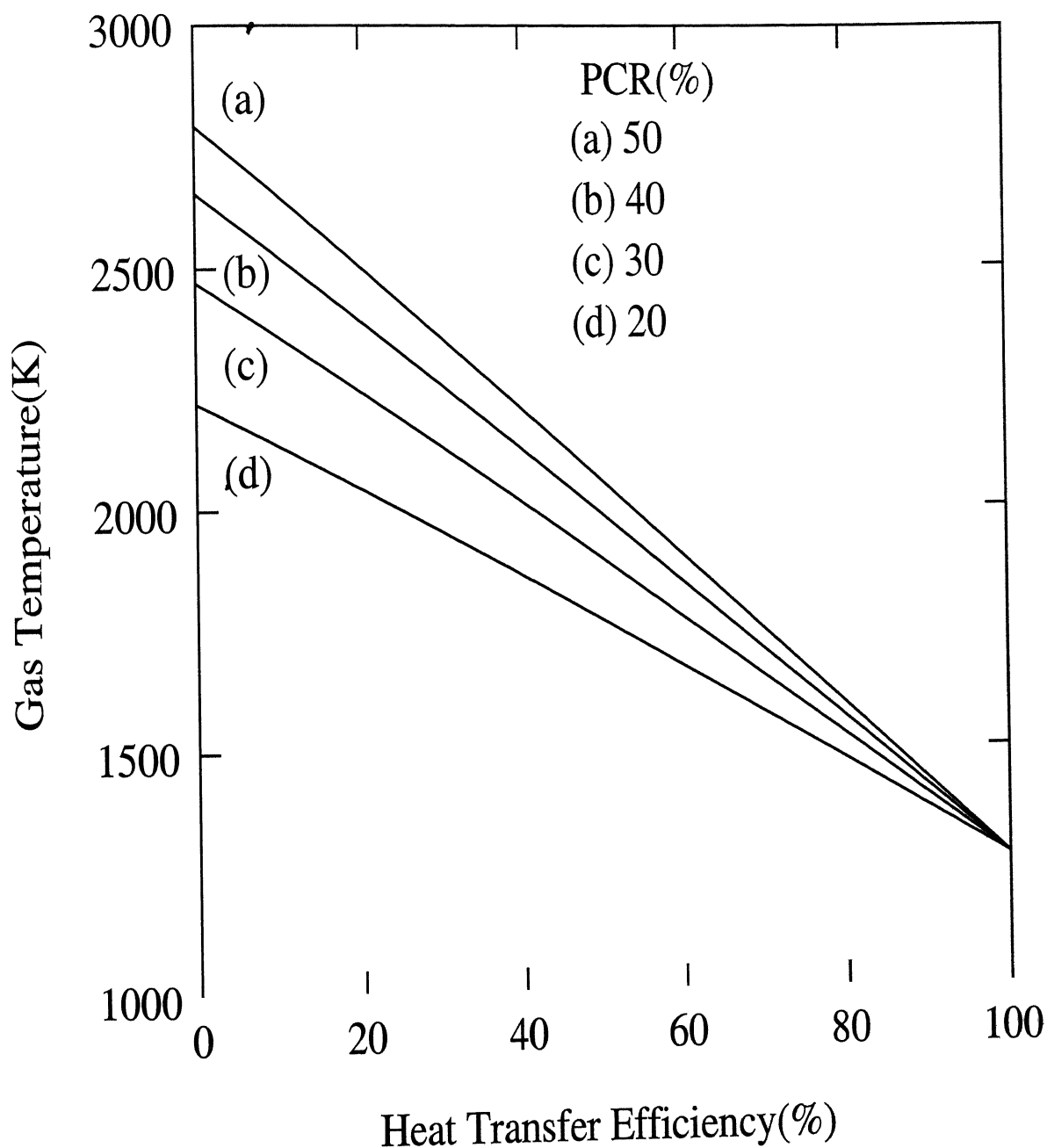


Figure 4.25: Gas temperature vs heat transfer efficiency at different post combustion ratio for 1000°K preheated air as post combustion medium

4.4.2 Coal rate

In figure 4.26 coal consumption has been plotted against heat transfer efficiency for different post combustion ratio. This figure is drawn by considering 30% metallization, bituminous coal composition and oxygen as post combustion medium. At 0% post combustion ratio, coal consumption rate does not vary with heat transfer efficiency because heat is not produced when there is no post combustion. But when we increase post combustion ratio to 10%, then some heat is produced for this post combustion of CO and H_2 . If the heat transfer to the bath is 70%, then coal consumption is 1113.4 Kg/thm and it decreases to 971.6 Kg/thm when heat transfer efficiency increases to 100% because heat supplied to the bath is more at high heat transfer efficiency. For 20% post combustion ratio, similarly coal consumption decreases from 856.3 Kg/thm to 700.4 Kg/thm (see table 4.2) when heat produced by this 20% post combustion is transferred to bath increases from 70% to 100%. So to reduce coal consumption heat transfer efficiency must be as high as possible. In this figure similarly for other post combustion ratio coal consumption decreases is shown.

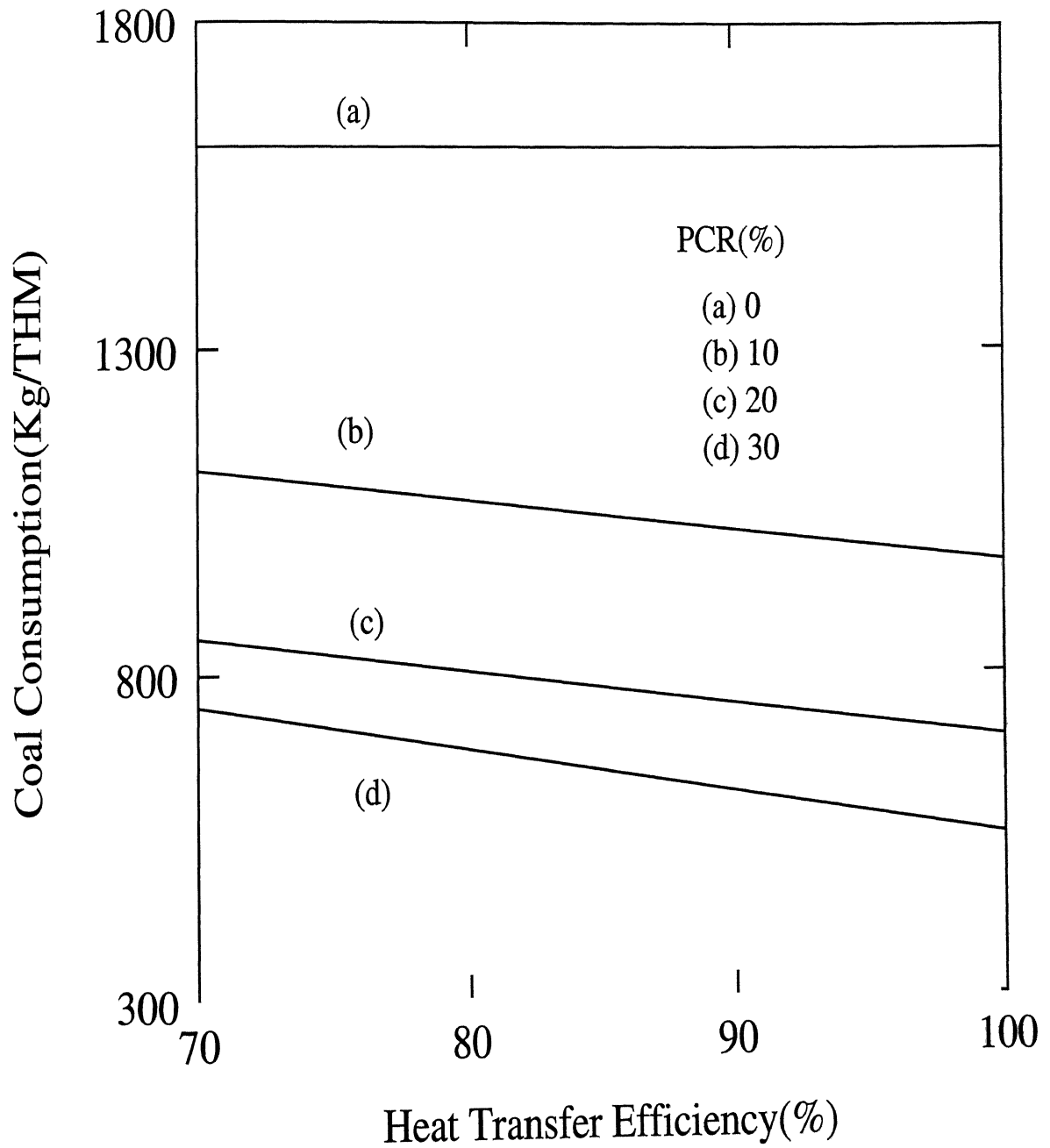


Figure 4.26: Coal consumption vs heat transfer efficiency at different post combustion ratio

Chapter 5

Conclusions

As a result of present study, the following conclusions are made :

(1) Increase in metallization decreases coal rate, flux rate, slag rate, oxygen requirement and export gas. At a given percent metallization, these are highest for subbituminous coal and lowest for anthracite coal.

(2) Post combustion greatly reduces coal consumption and hence flux rate, slag rate, oxygen required for coal. Post combustion reduces export gas also. However, post combustion ratio and degree of metallization must be chosen in such a way that the gas produced in the melter-gasifier is sufficient for reduction desired without reforming and scrubbing.

(3) One major problem is that the gas temperature goes too high to handle when post combustion ratio is very high. It is possible to reduce the temperature of gas by improving heat transfer to the bath or, using O_2 -air mixture for post combustion.

(4) Heat transfer must be above 71%, 67%, 60% and 44% when 50%, 40%, 30% and 20% post combustion is done by using oxygen and allowable temperature is $2000^\circ K$.

(5) At higher post combustion ratio, CO_2 dissociation occurs. When oxygen at $298^\circ K$ is used for post combustion, then dissociation starts at 37% post combustion. Instead of this dissociation after 37% post combustion ratio, the temperature of the gas increases with post combustion ratio because of increasing H_2 consumption for post combustion.

(6) Coal saving is highest when heat transfer efficiency and post combustion ratio is

high. So, heat transfer to the bath and post combustion ratio must be controlled to reduce coal consumption.

(7) Volume of the gas in the melter-gasifier decreases with increasing fraction of oxygen in the post combustion medium and hence increases gas temperature. So, fraction of oxygen in O_2 -air mixture for post combustion must be controlled.



Chapter 6

Suggestions For Further Work

(1) In the present study, the losses of heat and materials in the scrubber and hot gas cyclone are not considered. Energy requirement for scrubbing has also not been included in this work. A thermo-chemical model for the overall process can be developed by incorporating above information.

(2) Heat produced due to post combustion is transferred to the bath either through metal droplet behavior or through slag foaming. This heat transfer mechanism is not entered in this present work. So, model can be extended by including these.

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APPENDIX I

Calculations of variables in terms of kg moles of the components/kg moles of product iron.

$$n_C^{HM} = \frac{p_C^{HM}}{12} \cdot \frac{56}{p_{Fe}^{HM}}$$

$$n_{Si}^{HM} = \frac{p_{Si}^{HM}}{28} \cdot \frac{56}{p_{Fe}^{HM}}$$

$$n_{Mn}^{HM} = \frac{p_{Mn}^{HM}}{55} \cdot \frac{56}{p_{Fe}^{HM}}$$

$$n_S^{HM} = \frac{p_S^{HM}}{32} \cdot \frac{56}{p_{Fe}^{HM}}$$

$$n_P^{HM} = \frac{p_P^{HM}}{31} \cdot \frac{56}{p_{Fe}^{HM}}$$

M_I kg iron ore required to produce one kg mole of product iron.

M_I kg iron ore gives Fe_2O_3 ,

$$= M_I \times \frac{p_{Fe_2O_3}^{ore}}{100} \times \frac{1}{160}$$

As one kg mole of Fe_2O_3 corresponds to two kg moles of Fe. So, kg moles of iron produced from this M_I kg iron ore,

$$= M_I \times \frac{p_{Fe_2O_3}^{ore}}{100} \times \frac{1}{160} \times 2$$

Our basis of calculation is per kg mole of product iron.

So,

$$M_I \times \frac{p_{Fe_2O_3}^{ore}}{100} \times \frac{1}{160} \times 2 = 1$$

So,

$$M_I = \frac{100 \times 160}{p_{Fe_2O_3}^{ore} \times 2}$$

APPENDIX II

Calculation Of Gas Components

(all are expressed in per kg mole of product iron)

Without post combustion CO, H_2 and N_2 produced n_C^A , $\frac{1}{2}n_H^o$ and $\frac{1}{2}n_N^{coal}$ kg moles respectively.

If ppc percent of above gases is post combusted in the melter-gasifier, then kg moles of above gas components in the melter-gasifier after post combustion are :

$$\text{Kg moles of CO}(n_{CO}^r) = n_C^A \cdot \left(1 - \frac{ppc}{100}\right)$$

$$\text{Kg moles of } CO_2(n_{CO_2}^r) = n_C^A \cdot \frac{ppc}{100}$$

$$\text{Kg moles of } H_2(n_{H_2}^r) = \frac{1}{2}n_H^o \cdot \left(1 - \frac{ppc}{100}\right)$$

$$\text{Kg moles of } H_2O(n_{H_2O}^r) = \frac{1}{2}n_H^o \cdot \frac{ppc}{100}$$

$$\text{Kg moles of } N_2(n_{N_2}^r) = \frac{1}{2}n_N^{coal} + \frac{1}{2} \left[\frac{\%N_2}{\%O_2} \right]_{med}$$

When post combustion is done by oxygen, then $\left[\frac{\%N_2}{\%O_2} \right]_{med}$ term is zero.

So, total gas produced in the melter-gasifier,

$$n_t^g = n_{CO}^r + n_{CO_2}^r + n_{H_2}^r + n_{H_2O}^r + n_{N_2}^r$$

Kg moles of CO required in the reduction shaft furnace,

$$n_{CO}^{req} = \left[\frac{X}{100} \right] \cdot \frac{\left[3.3 - 1.22 [n_{H_2}^r]^n \right]}{0.95}$$

Kg moles of H_2 required in the reduction shaft furnace,

$$n_{H_2}^{req} = \frac{X}{100} \cdot \frac{[n_{H_2}^r]^n}{0.95}$$

Percentages of CO in the melter-gasifier,

$$p_{CO}^r = \frac{n_{CO}^r}{n_t^g} \times 100$$

Similarly $p_{CO_2}^r$, $p_{H_2}^r$, $p_{H_2O}^r$ and $p_{N_2}^r$ have been calculated.

kg moles of CO goes in the reduction shaft furnace = n_{CO}^{req}

kg moles of H_2 goes to the reduction shaft furnace = $n_{H_2}^{req}$

If, n_t^{rg} kg moles of gas goes to the reduction shaft furnace for reduction of iron ore.

So, surplus gas,

$$n_t^{sg} = n_t^g - n_t^{rg}$$

CO in the surplus gas,

$$n_{CO}^{sg} = n_t^{sg} \times \frac{p_{CO}^r}{100}$$

Similarly $n_{CO_2}^{sg}$, $n_{H_2}^{sg}$, $n_{H_2O}^{sg}$ and $n_{N_2}^{sg}$ have been calculated.

Now, utilization of CO and H_2 at $1200^\circ K$ for $Fe_{0.95}O - Fe$ equilibrium is 30% and 37% respectively.

So, top gas compositions are :

$$n_{CO}^{tg} = n_{CO}^{req} \times \frac{2.3}{3.3}$$

$$n_{H_2}^{tg} = n_{H_2}^{req} \times \frac{1.7}{2.7}$$

$$n_{CO_2}^{tg} = \frac{n_{CO}^{req}}{3.3}$$

$$n_{H_2O}^{tg} = \frac{n_{H_2}^{req}}{2.7}$$

$$n_{N_2}^{tg} = n_{N_2}^{rg}$$

Total top gas amount,

$$n_t^{tg} = n_{CO}^{tg} + n_{CO_2}^{tg} + n_{H_2}^{tg} + n_{H_2O}^{tg} + n_{N_2}^{tg}$$

APPENDIX III

Enthalpies of formation at temperature T from elements[16]

Compound	298°K	1100°K	1200°K	1300°K
CH_4	-75000	-91000	-91000	-92000
C_2H_4	-84000	-102000	-101000	-100000
C_3H_8	-105000	-129000	-128000	-127000
CO	-111000	-113000	-113000	-114000
CO_2	-394000	-395000	-395000	-395000
$CaCO_3$ (from CaO, C and O_2)	-572000	-563000	-561000	-559000
$CaCO_3 \cdot MgCO_3$ (from carbonates)	-12000	decomposes below 1100k		
$(CaO)_3 \cdot P_2O_5$ (from CaO, O_2 and P_{2g})	-2400000	-2364000	-2353000	-2341000
$CaOSiO_2$ (from oxides)	-89000	-91000	-91000	-91000
$(CaO)_2 \cdot SiO_2$	-126000	-125000	-124000	-123000
$Fe_{0.947}O$	-266000	-265000	-265000	-265000
Fe_3O_4	-1121000	-1096000	-1098000	-1095000
Fe_2O_3	-826000	-810000	-811000	-809000
H_2O_l ($\Delta H_{298}^{vaporization} = +44000$)	286000			
H_2O_g	-242000	-248000	-249000	-250000
$MgCO_3$ (from MgO, C and O_2)	-511000	decomposes below 1100°K		
MnO	-385000	-386000	-386000	-386000
Mn_3O_4	-1387000	-1387000	-1386000	-1384000
MnO_2	-521000	-516000	decomposes at 1120°K	
SO_2 (from gaseous S_2)	-362000	-362000	-362000	-362000
SiO_2	-911000	-902000	-901000	-901000

APPENDIX IV

Enthalpy increment equations for elements and compounds [$H_T^\circ - H_{298}^\circ$] [16]

Substance	298-800 Top gas	800-1100	1100-1300 Wustite reduction	1100-1900 Bosh & Hearth	1800-2800 Flame temperature
Al_2O_3	107T-32000	124T-46000	128T-50000	132T-55000	-
C	15.3T-4600	20.7T-8700	22.7T-11000	23.5T-118000	-
CO	30.2T-9100	32.5T-10800	34.2T-12600	35.3T-14000	36.6T-16400
CO_2	45.6T-141000	52.8T-19200	56.3T-23100	58.6T-26000	61.0T-30500
CaO	48.5T-14500	51.6T-17000	54.2T-19800	55.5T-212000	-
$CaCO_3$	106T-31500	123T-45100	129T-52200	Decomposes above 1300°K	
$CaO.SiO_2$	108T-32300	122T-43600	128T-49500	128T-50000	-
$(CaO)_2.SiO_2$	158T-47000	186T-69000	189T-73000	218T-105000	-
$(CaO)_3.P_2O_5$	284T-84500	356T-142000	402T-193000	403T-194000	-
Fe_s	30.8T-9200	48.9T-23700		38.4T-11400	-
Fe_l	-	-	-	44.0T-5800	-
$Fe_{0.947}O$	52.2T-15600	56.4T-18900	58.6T-21400	61.1T-24100	-
Fe_2O_3	135T-40200	158T-58800	142T-40600	144T-43000	-
Fe_3O_4	202T-60000	227T-58800	201T-53200	201T-53200	-
H_2	29.3T-8800	29.9T-9200	30.9T-10300	32.5T-12200	35.2T-17200
$H_2O_{(g)}$	35.8T-10800	40.0T-13800	43.7T-17900	47.4T-22200	-
MgO	45.6T-13600	50.9T-17800	52.2T-19300	55.0T-22300	-
$MgCO_3$	102T-30500	126T-49500	Decomposes below 1100°K		
Mn_s	31.0T-9200	44.3T-19900	38.2T-13200	48.6T-24600	-
Mn_l	-	-	-	48.6T-24600	-
MnO	49.2T-14700	53.7T-18300	56.9T-21800	57.1T-22000	-
MnO_2	68.2T-20300	77.3T-27600	-	-	-
Mn_3O_4	167T-49600	182T-61900	199T-80200	238T-123000	-
N_2	30.0T-9000	32.1T-10500	-	34.4T-13000	36.4T-16500
O_2	31.6T-9600	34.3T-11500	-	36.2T-13500	-
P_{2g}	34.9T-10500	36.7T-11900	37.1T-12300	37.3T-12600	-
S_{2g}	35.2T-10600	36.9T-11900	37.2T-12300	37.4T-12500	-
Si_s	23.6T-7000	26.4T-9300	27.2T-10200	27.1T-10100	-
Si_l	-	-	-	27.2T+40400	-
SiO_2	61.1T-18200	71.0T-25900	71.4T-24300	72.8T-26200	-

APPENDIX V

All are expressed per kg mole product iron

$$\text{Let, } \frac{n_{Si}^a}{n_{SiO_2}^b} = \alpha, \quad \frac{n_{Si}^c}{n_{SiO_2}^d} = \beta \quad \text{and} \quad \frac{n_{Si}^f}{n_{SiO_2}^g} = \phi$$

Where,

$$n_{Si}^a + n_{SiO_2}^b = n_{SiO_2}^{ore}$$

$$n_{Si}^c + n_{SiO_2}^d = n_{SiO_2}^{ash}$$

$$n_{Si}^f + n_{SiO_2}^g = n_{SiO_2}^L + n_{SiO_2}^D$$

So,

$$n_{Si}^a = \frac{\alpha}{\alpha+1} \cdot n_{SiO_2}^{ore}$$

$$n_{SiO_2}^b = \frac{1}{\alpha+1} \cdot n_{SiO_2}^{ore}$$

$$n_{Si}^c = \frac{\beta}{\beta+1} \cdot n_{SiO_2}^{ash}$$

$$n_{SiO_2}^d = \frac{1}{\beta+1} \cdot n_{SiO_2}^{ash}$$

$$n_{Si}^f = \frac{\phi}{\phi+1} \cdot n_{SiO_2}^L + n_{SiO_2}^D$$

$$n_{SiO_2}^g = \frac{1}{\phi+1} \cdot n_{SiO_2}^L + n_{SiO_2}^D$$

$$\text{Let, } \frac{n_S^a}{n_{CaS}^b} = \gamma \quad \text{and} \quad \frac{n_S^c}{n_{CaS}^d} = \delta$$

Where,

$$n_S^a + n_{CaS}^b = n_{FeS}^{ore}$$

$$n_S^c + n_{CaS}^d = n_S^{coal}$$

So,

$$n_S^a = \frac{\gamma}{\gamma+1} \cdot n_{FeS}^{ore}$$

$$n_{CaS}^b = \frac{1}{\gamma+1} \cdot n_{FeS}^{ore}$$

Similarly,

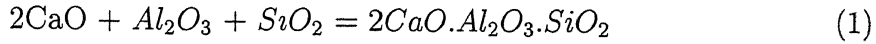
$$n_S^c = \frac{\delta}{\delta+1} \cdot n_S^{coal}$$

$$n_{CaS}^d = \frac{1}{\delta+1} \cdot n_S^{coal}$$

APPENDIX VI

Heat of slag formation

Reactions occurring in the slag are :



From reaction (1),

Kg moles of $2\text{CaO}.\text{Al}_2\text{O}_3.\text{SiO}_2 = \text{Kg moles of } \text{Al}_2\text{O}_3 \text{ in slag}(n_{\text{Al}_2\text{O}_3}^{\text{slag}})$

From reaction (4),

Kg moles of $\text{MnO}.\text{SiO}_2 = \text{Kg moles of MnO in slag}(n_{\text{MnO}}^{\text{slag}})$

From reaction (5),

Kg moles of $2\text{MgO}.\text{SiO}_2 = 2(\text{Kg moles of MgO in slag})(2n_{\text{MgO}}^{\text{slag}})$

Kg moles of $2\text{CaO}.\text{SiO}_2$ and $\text{CaO}.\text{SiO}_2$ are calculated as follows :

Let, kg moles of $2\text{CaO}.\text{SiO}_2 = a$

and kg moles of $\text{CaO}.\text{SiO}_2 = b$

From reactions (2) and (3),

Kg moles of CaO left for reaction (2) and (3),

$$2a + b = n_{\text{CaO}}^{\text{slag}} - 2n_{\text{Al}_2\text{O}_3}^{\text{slag}} \quad (6)$$

Kg moles of SiO_2 left for reaction (2) and (3) after reaction (1), (4) and (5),

$$a + b = n_{\text{SiO}_2}^{\text{slag}} - n_{\text{Al}_2\text{O}_3}^{\text{slag}} - n_{\text{MnO}}^{\text{slag}} - 0.5n_{\text{MgO}}^{\text{slag}} \quad (7)$$

By solving equation (6) and (7), we get,

Kg moles of $2\text{CaO}.\text{SiO}_2(a)$

$$= n_{\text{CaO}}^{\text{slag}} - n_{\text{Al}_2\text{O}_3}^{\text{slag}} - n_{\text{SiO}_2}^{\text{slag}} + n_{\text{MnO}}^{\text{slag}} + 0.5n_{\text{MgO}}^{\text{slag}}$$

and,

Kg moles of $\text{CaO}.\text{SiO}_2(b)$

$$= 2n_{SiO_2}^{slag} - 2n_{MnO}^{slag} - n_{CaO}^{slag} - n_{MgO}^{slag}$$

So, Heat produced in slag formation per kg moles of product Fe

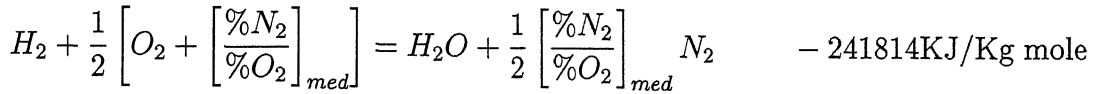
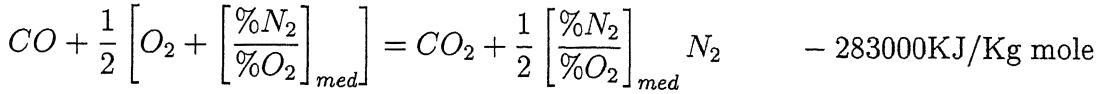
$$\begin{aligned} &= - (\text{Kg moles of } 2CaO.SiO_2).126840 - (\text{Kg moles of } CaO.SiO_2).90300 \\ &- (\text{Kg moles of } 2CaO.Al_2O_3.SiO_2).104580 - (\text{Kg moles of } MnO.SiO_2).24780 \\ &- (\text{Kg moles of } MgO.SiO_2).63420 \end{aligned}$$

$$\begin{aligned} &= - (n_{CaO}^{slag} - n_{Al_2O_3}^{slag} - n_{SiO_2}^{slag} + n_{MnO}^{slag} + 0.5n_{MgO}^{slag}).126840 \\ &- (2n_{SiO_2}^{slag} - 2n_{MnO}^{slag} - n_{CaO}^{slag} - n_{MgO}^{slag}).90300 \\ &- n_{Al_2O_3}^{slag}.104580 - n_{MnO}^{slag}.24780 - 2n_{MgO}^{slag}.63420 \end{aligned}$$

APPENDIX VII

Gas temperature and dissociation calculation

The post combustion of CO and H_2 occur by these two reactions :



Gas temperature is calculated as follows :

Sensible heat of the gas before post combustion + sensible heat of the post combustion medium + heat left in the post combustion zone = sensible heat of the combustion product gases (a)

Before post combustion gas contains CO, H_2 and N_2 .

So, sensible heat of the gases before post combustion at $1273^\circ K$,

$$\Delta H_1 = n_C^A \cdot \int_{298}^{1273} C_{p,CO} dT + \frac{1}{2} n_H^o \cdot \int_{298}^{1273} C_{p,H_2} dT + \frac{1}{2} n_N^{coal} \cdot \int_{298}^{1273} C_{p,N_2} dT \quad (b)$$

O_2 comes from post combustion medium = $\frac{1}{2} (n_{CO_2}^r + n_{H_2O}^r)$

and N_2 comes from post combustion medium = $\frac{1}{2} \left[\frac{\%N_2}{\%O_2} \right]_{med} (n_{CO_2}^r + n_{H_2O}^r)$

So, sensible heat of post combustion medium at $T_p^\circ K$,

$$\Delta H_2 = \frac{1}{2} (n_{CO_2}^r + n_{H_2O}^r) \cdot \int_{298}^{T_p} C_{p,O_2} dT + \frac{1}{2} \left[\frac{\%N_2}{\%O_2} \right]_{med} \cdot (n_{CO_2}^r + n_{H_2O}^r) \cdot \int_{298}^{T_p} C_{p,N_2} dT \quad (c)$$

Heat left in the post combustion zone,

$$\Delta H_3 = \left[(n_{CO_2}^r - \epsilon) \times 283000 + (n_{H_2O}^r + \epsilon) \times 241814 \right] \cdot \left[1 - \frac{HTE}{100} \right] \quad (d)$$

After post combustion gas contains CO, CO_2 , H_2 , H_2O and N_2 . So, after post combustion,

$$\text{Kg moles of CO} = (n_{CO}^r + \epsilon)$$

$$\text{Kg moles of } CO_2 = (n_{CO_2}^r - \epsilon)$$

$$\text{Kg moles of } H_2 = (n_{H_2}^r - \epsilon)$$

$$\text{Kg moles of } H_2O = (n_{H_2O}^r + \epsilon)$$

$$\text{Kg moles of } N_2 = \frac{1}{2} \left[n_N^{coal} + \left[\frac{\%N_2}{\%O_2} \right]_{med} \cdot [n_{CO_2}^r + n_{H_2O}^r] \right]$$

So, sensible heat of the post combustion product gases at $T_g^\circ K$,

$$\begin{aligned} \Delta H_4 = & (n_{CO}^r + \epsilon) \cdot \int_{298}^{T_g} C_{p,CO} dT + (n_{CO_2}^r - \epsilon) \cdot \int_{298}^{T_g} C_{p,CO_2} dT + (n_{H_2}^r - \epsilon) \cdot \int_{298}^{T_g} C_{p,H_2} dT \\ & + (n_{H_2O}^r + \epsilon) \cdot \int_{298}^{T_g} C_{p,H_2O} dT + \frac{1}{2} \left[n_N^{coal} + \left[\frac{\%N_2}{\%O_2} \right]_{med} \cdot [n_{CO_2}^r + n_{H_2O}^r] \right] \cdot \int_{298}^{T_g} C_{p,N_2} dT \quad (e) \end{aligned}$$

combining equation (a) - (e), we get,

$$\begin{aligned} & n_C^A \cdot \int_{298}^{1273} C_{p,CO} dT + \frac{1}{2} n_H^O \cdot \int_{298}^{1273} C_{p,H_2} dT + \frac{1}{2} n_N^{coal} \cdot \int_{298}^{1273} C_{p,N_2} dT \\ & + \frac{1}{2} (n_{CO_2}^r + n_{H_2O}^r) \cdot \int_{298}^{T_p} C_{p,O_2} dT + \frac{1}{2} \left[\frac{\%N_2}{\%O_2} \right]_{med} \cdot (n_{CO_2}^r + n_{H_2O}^r) \cdot \int_{298}^{T_p} C_{p,N_2} dT \\ & + \left[(n_{CO_2}^r - \epsilon) \times 283000 + (n_{H_2O}^r + \epsilon) \times 241814 \right] \cdot \left[1 - \frac{HTE}{100} \right] \\ & = (n_{CO}^r + \epsilon) \cdot \int_{298}^{T_g} C_{p,CO} dT + (n_{CO_2}^r - \epsilon) \cdot \int_{298}^{T_g} C_{p,CO_2} dT + (n_{H_2}^r - \epsilon) \cdot \int_{298}^{T_g} C_{p,H_2} dT \\ & + (n_{H_2O}^r + \epsilon) \cdot \int_{298}^{T_g} C_{p,H_2O} dT + \frac{1}{2} \left[n_N^{coal} + \left[\frac{\%N_2}{\%O_2} \right]_{med} \cdot [n_{CO_2}^r + n_{H_2O}^r] \right] \cdot \int_{298}^{T_g} C_{p,N_2} dT \quad (f) \end{aligned}$$

In this equation kg moles of CO_2 dissociation ϵ is unknown. Gas temperature T_g has been calculated from this equation after calculating ϵ

Calculation of ϵ is as follows :

Before dissociation at higher post combustion ratio the temperature of the gas increases too high and dissociation starts and after that the temperature of the gas decreases to T_g which has been calculated from equation (f).

Now, to know the degree of dissociation, first the temperature of the gas before dissociation (T_f) has been calculated by considering before dissociation i.e. $\epsilon = 0$ and putting T_f instead of T_g in equation (f).

Now, during dissociation these two reactions occurs :



Free energy change for reaction (g),

$$\begin{aligned} \Delta G_1^0 &= \Delta G_{f,CO}^0 - \Delta G_{f,CO_2}^0 \\ &= (-112877 - 86.514T) - (-394762 - 0.836T) \\ &= 281885 - 85.678T \end{aligned}$$

Where, $\Delta G_{f,CO}^0$ = standard free energy of formation of CO

and $\Delta G_{f,CO_2}^0$ = standard free energy of formation of CO_2

At equilibrium, $\Delta G_1^0 = 0$

So, from above equation temperature (T) of dissociation of CO_2 starts are calculated.

Similarly, for H_2O dissociation, free energy change for reaction (h),

$$\Delta G_2^0 = -247392 + 55.849T$$

At equilibrium, $\Delta G_2^0 = 0$

So, from above equation T is calculated. Above this temperature ΔG_2^0 value becomes positive, hence opposite reaction i.e. H_2O starts to dissociate.

Now, for reaction (g), equilibrium constant

$$K_1 = \exp \left(\frac{\Delta G_1^0}{-RT_f} \right) \quad (i)$$

and, for reaction (h), equilibrium constant,

$$K_2 = \exp \left(\frac{\Delta G_2^0}{-RT_f} \right) \quad (j)$$

From (i) and (j), K_1 and K_2 can be known.

Now, total kg moles of gas after dissociation,

$$TOTAL = n_{CO}^r + n_{CO_2}^r + n_{H_2}^r + n_{H_2O}^r + \frac{1}{2} \left[n_N^{coal} + \left[\frac{\%N_2}{\%O_2} \right]_{med} \cdot [n_{CO_2}^r + n_{H_2O}^r] \right]$$

So, partial pressure of CO in the gas,

$$p_{CO} = \frac{n_{CO}^r + \epsilon}{TOTAL}$$

Similarly, partial pressure of CO_2 , H_2 , H_2O and N_2 are calculated.

Now, if we write equilibrium constant of reaction (g) and (h) in terms of partial pressure,

$$K_1 = \frac{p_{CO} \cdot \sqrt{p_{O_2}}}{p_{CO_2}} \quad (k)$$

and

$$K_2 = \frac{p_{H_2O}}{p_{H_2} \cdot \sqrt{p_{O_2}}} \quad (l)$$

Multiplying equations (k) and (l), we get

$$\begin{aligned} K_1 \cdot K_2 &= \frac{p_{CO} \cdot p_{H_2O}}{p_{CO_2} \cdot p_{H_2}} \\ &= \frac{(n_{CO_2}^r + \epsilon) \cdot (n_{H_2O}^r + \epsilon)}{(n_{CO_2}^r - \epsilon) \cdot (n_{H_2}^r - \epsilon)} \quad (m) \end{aligned}$$

In this equation we know the value of K_1 , K_2 . So, we can calculate ϵ from this equation (m).

APPENDIX VIII

Compositions of different coal(dry basis)[17]

Group (ASTM designation D 388)	Ultimate analyses(%)					
	Carbon	Hydrogen	Nitrogen	Sulpher	Oxygen	Ash
Metaanthracite	86.8	1.6	0.6	0.9	2.0	8.1
Anthracite	84.2	2.8	0.8	0.6	2.2	9.4
Semianthracite	78.3	3.6	1.4	2.0	2.3	12.4
Bituminous						
Low-volatile	85.4	4.8	1.5	0.8	2.6	4.9
Medium-volatile	86.4	4.9	1.6	0.6	3.6	2.9
High-volatile A	79.5	5.2	1.4	1.3	6.1	6.5
High-volatile B	78.3	5.2	1.5	1.4	8.2	5.4
High-volatile C	73.1	4.8	1.5	2.6	8.9	9.1
Subbituminous						
A	75.1	5.0	1.4	1.0	12.8	4.7
B	75.0	4.9	1.3	0.5	15.5	2.8
C	64.2	4.4	1.2	0.4	16.6	13.2

APPENDIX IX

Compositions of different coal(dry ash free basis) with calorific values[17]

Group (ASTM designation D 388)	Volatile matter	Fixed carbon	Gross caloric value(Btu/lb)
Metaanthracite	1.3	98.7	31879
Anthracite	3.7	36.3	32177
Semianthracite	14.8	85.2	31641
Bituminous			
Low-volatile	16.8	83.2	34950
Medium-volatile	22.8	77.2	35368
High-volatile A	36.7	63.3	33453
High-volatile B	41.4	58.6	32872
High-volatile C	40.0	60.0	31383
Subbituminous			
A	40.8	59.2	31676
B	44.0	56.0	30868
C	44.4	54.6	25863

APPENDIX X

Table 4.1 : Heat demand for various metallurgical operations in melter-gasifier at different metallization

Metallization (%)	0	20	40	60	80	100
Component						
Hot metal	0.9233	0.9233	0.9233	0.9233	0.9233	0.9233
Slag	0.1372	0.1221	0.1059	0.0908	0.0756	0.0605
Heat absorbed by endothermic reactions	7.157	6.254	5.3906	4.527	3.664	2.8393
Sensible heat of the reduction gas leaving MG	5.832	5.1026	4.3626	3.6329	2.9033	2.1737
Drying of coal	0.0388	0.034	0.0291	0.0243	0.0195	0.0147
Latent heat of devolatilization of coal	1.2378	1.0846	0.9294	0.7762	0.6231	0.4699
Heat losses	0.3382	0.3382	0.3382	0.3382	0.3382	0.3382

* All are in GJ/THM.

Table 4.2 : Coal rate at various degree of metallization for different post combustion ratio for bituminus and subbituminus coal when pure oxygen is used for post combustion

Heat transfer efficiency	PCR		0	10	20	30	40	50
	% M							
70	0	B	1977.2	1367.4	1051.6	922.8	731.1	657.5
		S	2440.1	1687.3	1297.5	1138.6	901.8	811.8
	30	B	1606.9	1113.4	856.3	751.0	595.1	535.5
		S	1985.6	1373.9	1056.7	926.2	734.3	661.1
	60	B	1239.9	858.1	659.7	578.6	458.6	412.5
		S	1530.3	1058.9	813.2	713.3	567.1	508.5
	90	B	873.3	604.5	466.1	407.6	323.2	290.7
		S	1078.1	746.2	574.9	503.2	399.1	359.3
80	0	B	1977.2	1307.4	979.4	786.5	664.1	593.6
		S	2440.1	1612.5	1201.2	970.6	818.1	733.1
	30	B	1606.9	1064.5	797.3	640.3	540.9	483.3
		S	1985.6	1314.1	981.2	789.3	666.6	597.1
	60	B	1239.9	819.9	615.1	494.2	418.1	373.3
		S	1530.3	1011.4	758.8	609.9	517.3	460.3
	90	B	873.3	577.8	432.8	347.5	293.1	263.5
		S	1078.1	712.7	533.8	428.3	361.7	324.7
90	0	B	1977.2	1248.5	915.5	727.1	610.2	541.1
		S	2440.1	1540.3	1129.4	897.4	752.7	667.7
	30	B	1606.9	1016.7	745.7	591.3	495.6	440.2
		S	1985.6	1254.3	919.9	729.4	611.3	542.9
	60	B	1239.9	784.3	576.1	456.9	383.4	340.3
		S	1530.3	967.4	711.3	563.8	473.1	419.7
	90	B	873.3	551.9	405.1	320.9	269.1	238.9
		S	1078.1	681.2	499.3	394.5	331.5	294.7
100	0	B	1977.2	1193.6	860.5	674.7	561.5	503.3
		S	2440.1	1474.3	1063.2	832.5	692.5	621.2
	30	B	1606.9	971.6	700.4	549.1	457.2	404.2
		S	1985.6	1196.2	861.9	676.3	563.7	498.3
	60	B	1239.9	749.2	539.7	424.3	352.9	305.4
		S	1530.3	925.1	665.3	525.1	436.7	376.5
	90	B	873.3	527.3	381.5	299.4	249.2	206.1
		S	1078.1	651.2	473.8	369.1	307.4	254.7

* B→ Bituminus coal, S→ Subbituminus coal

Table 4.3 : Coal rate at various degree of metallization for different post combustion ratio, heat transfer efficiency for bituminus and subbituminus coal when 40% O₂ enriched air is used for post combustion

Heat transfer efficiency	PCR		0	10	20	30	40	50
	% M							
70	0	B	1977.2	1423.2	1115.9	920.1	783.3	684.2
		S	2440.1	1750.3	1375.1	1130.5	961.7	843.5
	30	B	1606.9	1160.9	911.7	752.1	641.7	651.1
		S	1985.6	1431.5	1120.5	926.1	788.1	691.3
	60	B	1239.9	896.5	705.9	584.1	500.1	438.0
		S	1530.3	1106.2	872.9	718.1	616.2	538.7
	90	B	873.3	634.3	501.8	416.1	357.4	314.9
		S	1078.1	780.2	620.2	511.7	439.2	384.2
80	0	B	1977.2	1354.8	1034.8	838.0	706.3	611.8
		S	2440.1	1655.2	1261.2	1022.3	857.4	746.0
	30	B	1606.9	1110.3	850.1	688.3	577.9	501.4
		S	1985.6	1354.3	1037.2	839.5	704.1	611.2
	60	B	1239.9	853.4	652.1	527.6	444.9	385.5
		S	1530.3	1039.4	794.3	642.6	541.6	469.1
	90	B	873.3	602.7	464.4	380.1	322.7	281.2
		S	1078.1	734.1	566.5	462.4	392.7	342.8
90	0	B	1977.2	1292.5	964.5	770.1	642.9	553.1
		S	2440.1	1575.6	1176.7	938.2	783.2	674.7
	30	B	1606.9	1059.4	789.9	631.2	526.3	453.5
		S	1985.6	1291.9	962.1	769.8	641.1	552.7
	60	B	1239.9	811.6	607.1	485.6	405.3	348.4
		S	1530.3	989.9	740.3	591.2	494.2	468.3
	90	B	873.3	575.9	433.5	384.7	293.3	254.5
		S	1078.1	703.2	529.1	469.7	357.2	308.5
100	0	B	1977.2	1238.9	906.7	716.6	593.3	512.4
		S	2440.1	1527.7	1116.8	882.3	730.5	630.6
	30	B	1606.9	1006.9	737.0	582.4	482.8	415.5
		S	1985.6	1241.2	907.7	717.5	595.1	512.5
	60	B	1239.9	774.7	567.3	447.7	372.2	318.6
		S	1530.3	955.7	699.1	552.4	458.3	394.8
	90	B	873.3	542.8	396.9	313.9	262.4	221.3
		S	1078.1	735.9	538.5	425.5	353.0	304.1

* B→ Bituminus coal, S→ Subbituminus coal

Table 4.4 : Coal rate at various degree of metallization for different post combustion ratio and heat transfer efficiency for bituminus and subbituminus coal when air is used for post combustion

Heat transfer efficiency	PCR		0	10	20	30	40	50
	% M							
70	0	B	1977.2	1504.5	1216.7	1023.8	883.9	779.2
		S	2440.1	1855.9	1501.2	1263.2	1090.2	961.2
	30	B	1606.9	1226.2	985.3	829.0	715.8	631.2
		S	1985.6	1512.3	1216.4	1024.5	882.4	778.7
	60	B	1239.9	948.1	698.3	642.7	554.9	489.7
		S	1530.3	1172.1	861.4	795.9	686.3	555.3
	90	B	873.3	670.3	544.4	461.8	401.9	356.6
		S	1078.1	826.8	672.5	567.2	492.4	435.6
80	0	B	1977.2	1426.4	1119.5	923.4	787.2	686.6
		S	2440.1	1754.8	1377.2	1135.3	371.4	844.1
	30	B	1606.9	1156.1	907.3	748.3	638.3	556.5
		S	1985.6	1426.5	1116.8	918.1	782.4	691.0
	60	B	1239.9	895.4	702.4	579.9	494.7	431.4
		S	1530.3	1099.5	865.0	708.3	610.8	531.8
	90	B	873.3	635.2	502.4	416.4	357.4	314.6
		S	1078.1	783.6	619.5	511.5	438.9	384.7
90	0	B	1977.2	1357.7	1023.2	824.5	690.1	593.5
		S	2440.1	1673.6	1260.9	1024.2	851.2	728.1
	30	B	1606.9	1102.5	824.6	696.3	589.8	512.5
		S	1985.6	1359.9	1010.2	851.9	722.8	631.8
	60	B	1239.9	847.3	625.8	568.1	489.5	431.5
		S	1530.3	1044.1	769.2	697.9	599.4	526.8
	90	B	873.3	592.1	427.1	439.8	389.2	350.5
		S	1078.1	729.6	524.9	538.7	476.2	425.6
100	0	B	1977.2	1269.8	923.6	721.4	593.8	501.4
		S	2440.1	1564.9	1136.9	886.7	727.7	613.5
	30	B	1606.9	1055.6	789.3	631.4	527.3	453.9
		S	1985.6	1300.1	971.5	777.3	644.5	550.3
	60	B	1239.9	841.4	655.1	541.3	460.8	406.4
		S	1530.3	1036.1	805.2	663.2	561.2	500.4
	90	B	873.3	627.2	520.7	451.3	394.3	358.9
		S	1078.1	771.7	638.4	551.5	485.4	430.7

* B→ Bituminus coal, S→ Subbituminus coal

Table 4.5 : O₂ requirement for coal at various degrees of metallization for different post combustion ratio and heat transfer efficiency for bituminus and subbituminus coal when pure oxygen is used for post combustion

Heat transfer efficiency	PCR		0	10	20	30	40	50
	% M							
70	0	B	3444.9	2410.4	1875.1	1546.7	1321.5	1165.3
		S	4271.7	2988.4	2326.3	1919.5	1634.3	1443.4
	30	B	2731.8	1896.7	1463.8	1194.9	1030.3	952.6
		S	3387.5	2351.0	1812.9	1478.1	1273.4	1178.4
	60	B	2022.8	1382.5	1052.5	843.1	739.1	659.9
		S	2507.6	1713.7	1301.5	1042.9	911.4	812.2
	90	B	1313.7	868.0	641.2	491.3	447.9	417.2
		S	1627.6	1072.6	793.6	606.4	551.4	513.6
80	0	B	3444.9	2292.1	738.8	1400.6	1194.8	1057.4
		S	4271.7	2840.8	2155.3	1795.9	1475.6	1305.7
	30	B	2731.8	1810.8	1373.1	1113.6	943.4	835.1
		S	3387.5	2243.2	1700.0	1376.4	1164.4	1029.2
	60	B	2022.8	1345.4	1020.2	827.7	700.9	620.5
		S	2507.6	1666.6	1262.3	1021.7	863.0	762.6
	90	B	1313.7	819.4	577.4	421.2	364.5	334.3
		S	1627.6	1014.3	713.0	518.3	446.4	407.9
90	0	B	3444.9	2173.8	1601.9	1273.4	1068.2	949.4
		S	4271.7	2693.3	1972.8	1574.8	1319.4	1170.6
	30	B	2731.8	1705.3	1244.2	837.4	759.3	715.1
		S	3387.5	2112.9	1540.1	1034.2	936.2	880.4
	60	B	2022.8	1237.2	869.2	605.2	526.5	483.4
		S	2507.6	1532.6	1075.1	746.5	647.3	483.4
	90	B	1313.7	769.1	514.4	371.6	294.5	251.7
		S	1627.6	952.3	634.9	458.3	359.6	305.0
100	0	B	3444.9	2055.5	1465.4	1136.8	941.5	841.5
		S	4241.6	2546.9	1814.1	1404.9	1161.9	1036.6
	30	B	2731.8	1610.2	1127.5	861.7	701.0	617.4
		S	3377.6	1991.2	1393.3	1064.5	833.3	762.9
	60	B	2022.8	1159.6	788.5	586.6	479.2	393.3
		S	2507.6	1435.9	974.6	722.9	589.1	481.1
	90	B	1313.7	719.6	451.2	311.5	239.4	169.2
		S	1627.6	890.3	556.8	381.9	291.5	209.6

* B→ Bituminus coal, S→ Subbituminus coal

Table 4.6 : O₂ requirement for coal at various degrees of metallization for different post combustion ratio and heat transfer efficiency for bituminous and subbituminous coal when 40% O₂-enriched air is used for post combustion

Heat transfer efficiency	PCR		0	10	20	30	40	50
	% M							
70	0	B	3444.9	2527.8	2020.3	1697.7	1471.7	1309.4
		S	4271.7	3158.7	2522.5	2117.5	1835.1	1630.2
	30	B	2731.8	1991.3	1580.3	1318.5	1135.9	1004.9
		S	3387.5	2487.5	1968.7	1637.2	1412.5	1247.3
	60	B	2022.8	1455.1	1140.6	739.4	800.0	700.3
		S	2507.6	1817.5	1418.8	913.8	993.8	864.3
	90	B	1313.7	919.6	701.4	560.2	464.2	395.8
		S	1627.6	1147.4	870.6	690.1	572.5	487.6
80	0	B	3444.9	2400.2	1866.9	1541.2	1319.1	1161.2
		S	4271.7	2998.8	2326.3	1991.6	1642.5	1441.2
	30	B	2731.8	1887.5	1455.9	1190.1	1010.9	883.2
		S	3387.5	2357.6	1812.4	1475.6	1256.3	1100.6
	60	B	2022.8	1374.3	1044.9	839.2	702.7	605.4
		S	2507.6	1716.3	1298.8	1038.8	871.3	748.8
	90	B	1313.7	861.6	633.9	488.3	394.5	327.5
		S	1627.6	1018.7	786.6	601.3	486.3	400.2
90	0	B	3444.9	2272.7	1713.6	1384.7	1166.5	1013.1
		S	4271.7	2838.7	2134.9	1726.3	1451.3	1256.1
	30	B	2731.8	1783.7	1331.6	1062.7	885.9	761.4
		S	3387.5	2227.5	1657.3	1315.2	1101.2	944.2
	60	B	2022.8	1294.7	931.5	740.6	605.6	509.7
		S	2507.6	1616.3	1156.3	915.9	751.3	621.9
	90	B	1313.7	805.6	531.4	418.1	324.7	258.1
		S	1627.6	1004.1	657.5	512.6	400.9	311.1
100	0	B	3444.9	2145.1	1560.2	1228.2	1013.9	864.8
		S	4271.7	2680.2	1943.8	1525.7	1260.3	1060.7
	30	B	2731.8	1679.9	1207.2	935.9	761.1	639.7
		S	3387.5	2079.3	1501.9	1158.8	935.7	778.4
	60	B	2022.8	1214.7	854.3	643.6	508.3	414.6
		S	2507.6	1514.3	1061.3	992.4	623.2	505.3
	90	B	1313.7	749.5	501.3	351.3	255.5	189.5
		S	1627.6	930.3	617.5	430.3	310.2	226.7

* B→ Bituminous coal, S→ Subbituminous coal

Table 4.7 : O₂ requirement for coal at various degrees of metallization for different post combustion ratio and heat transfer efficiency for bituminous and subbituminous coal when air is used for post combustion

Heat transfer efficiency	PCR		0	10	20	30	40	50
	% M							
70	0	B	3444.9	2693.8	2236.9	1931.7	1709.9	1544.5
		S	4271.7	3097.9	2572.4	2220.7	1963.1	1772.2
	30	B	2731.8	2086.2	1755.6	1508.9	1329.9	1195.9
		S	3387.5	2398.9	2017.1	1731.9	1523.8	1368.5
	60	B	2022.8	1539.5	1274.2	1086.2	949.8	847.4
		S	2507.6	1769.9	1401.4	1172.9	1005.9	901.2
	90	B	1313.7	992.9	792.9	663.5	569.8	498.8
		S	1627.6	1140.2	871.2	716.9	603.1	527.6
80	0	B	3444.9	2551.4	2058.1	1742.3	1521.5	1358.9
		S	4271.7	2930.4	2270.2	1881.6	1621.9	1448.5
	30	B	2731.8	2011.0	1610.3	1355.2	1175.8	1043.9
		S	3387.5	2310.5	1773.5	1463.7	1249.4	1125.7
	60	B	2022.8	1470.7	1162.4	967.7	830.1	729.0
		S	2507.6	1685.7	1278.5	1044.8	880.4	792.8
	90	B	1313.7	930.3	714.6	580.4	484.4	414.1
		S	1627.6	1065.4	785.7	627.8	513.1	459.1
90	0	B	3444.9	2409.1	1879.3	1552.8	1332.1	1173.4
		S	4271.7	2770.4	2066.4	1676.1	1411.9	1269.2
	30	B	2731.8	1895.3	1465.3	1200.7	1021.0	892.1
		S	3387.5	2175.9	1611.9	1301.4	1090.4	957.3
	60	B	2022.8	1381.4	1050.7	848.6	709.9	610.7
		S	2507.6	1588.2	1155.2	918.0	755.4	662.9
	90	B	1313.7	867.6	636.4	496.5	398.9	329.4
		S	1627.6	997.0	699.7	535.7	423.7	363.2
100	0	B	3444.9	2266.6	1700.5	1363.4	1143.2	987.8
		S	4271.7	2601.3	1875.6	1480.5	1211.9	1058.3
	30	B	2731.8	1779.4	1319.7	1046.6	866.6	740.1
		S	3387.5	2046.4	1445.4	1129.3	920.9	795.1
	60	B	2022.8	1292.2	938.9	729.8	590.0	492.4
		S	2507.6	1487.3	1031.8	792.4	627.9	532.4
	90	B	1313.7	805.0	558.1	413.0	313.4	244.7
		S	1627.6	930.7	614.9	450.1	345.5	262.8

* B→ Bituminous coal, S→ Subbituminous coal

Table 4.9 : Slag produced at different degrees of metallization for different post combustion ratio and heat transfer efficiency for bituminous and subbituminous coal

Heat transfer efficiency	PCR		0	10	20	30	40	50
	% M							
70	0	B	308.7	268.9	249.7	237.9	228.1	223.3
		S	394.3	341.8	316.2	299.5	291.4	279.1
	30	B	284.1	253.6	238.0	227.4	220.3	216.0
		S	349.2	322.6	302.1	286.7	276.5	270.1
	60	B	261.3	238.3	226.3	216.9	212.5	208.7
		S	334.1	303.4	286.7	272.6	266.3	259.8
	90	B	238.6	223.2	214.6	206.9	204.7	201.4
		S	304.6	284.2	271.4	263.7	256.3	250.9
	0	B	308.7	265.2	245.8	234.7	225.1	219.8
		S	394.3	337.9	311.0	295.7	276.5	262.5
	30	B	284.1	249.3	234.6	225.2	217.9	213.4
		S	349.2	317.4	299.3	284.2	272.6	261.9
80	60	B	261.3	235.7	223.3	215.6	210.8	207.0
		S	334.1	300.8	285.4	271.4	268.8	254.6
	90	B	298.6	221.0	212.0	206.1	203.6	200.6
		S	304.6	281.6	268.4	259.1	256.6	249.5
	0	B	308.7	261.6	241.9	230.9	222.2	216.7
		S	394.3	331.5	305.9	291.6	273.4	265.7
	30	B	284.1	247.3	231.1	222.7	215.9	210.9
		S	349.2	314.9	292.5	280.5	264.5	258.3
	60	B	261.3	233.1	220.3	213.7	209.1	205.4
		S	334.1	295.9	278.6	267.7	257.1	252.2
	90	B	238.6	218.8	209.5	205.1	202.4	199.9
		S	304.6	276.5	264.9	257.6	248.5	249.1
90	0	B	308.7	257.9	238.1	226.6	219.2	212.9
		S	374.3	308.5	286.3	273.6	262.5	257.4
	30	B	284.1	245.4	227.5	219.1	212.3	208.3
		S	349.2	300.1	278.1	268.3	259.6	254.7
	60	B	261.3	231.0	216.9	211.6	206.4	203.7
		S	334.1	293.2	269.3	262.1	255.4	251.2
	90	B	238.6	216.6	206.9	204.1	201.3	199.1
		S	304.6	277.3	261.1	257.3	252.2	248.3
	0	B	308.7	257.9	238.1	226.6	219.2	212.9
		S	374.3	308.5	286.3	273.6	262.5	257.4
	30	B	284.1	245.4	227.5	219.1	212.3	208.3
		S	349.2	300.1	278.1	268.3	259.6	254.7
100	60	B	261.3	231.0	216.9	211.6	206.4	203.7
		S	334.1	293.2	269.3	262.1	255.4	251.2
	90	B	238.6	216.6	206.9	204.1	201.3	199.1
		S	304.6	277.3	261.1	257.3	252.2	248.3

* B→ Bituminous coal, S→ Subbituminous coal

Table 4.10: Surplus gas produced at different degrees of metallization for different post combustion ratio and heat transfer efficiency for bituminous and subbituminous coal when pure oxygen is used for post combustion

Heat transfer efficiency	PCR		0	10	20	30	40	50
	% M							
70	0	B	4058.0	2205.6	1270.3	728.3	389.9	221.1
		S	5058.5	2772.0	1611.4	939.9	520.3	311.9
	30	B	3074.0	1591.0	867.3	462.9	225.0	84.0
		S	3976.8	2032.0	1135.5	632.4	337.5	162.0
	60	B	1914.9	820.9	308.1	43.2	0	0
		S	2557.8	1137.7	505.6	171.0	0	0
	90	B	755.8	52.7	0	0	0	0
		S	1230.2	243.3	0	0	0	0
80	0	B	4058.0	2105.1	1193.5	684.3	371.5	206.3
		S	5058.5	2653.1	1520.5	888.2	500.2	293.1
	30	B	3074.0	1499.8	785.4	401.9	182.9	56.9
		S	3976.8	1923.2	1039.6	559.4	286.6	139.7
	60	B	1914.9	750.4	246.5	0	0	0
		S	2557.8	1039.5	404.4	0	0	0
	90	B	755.8	0.9	0	0	0	0
		S	1230.2	155.5	0	0	0	0
90	0	B	4058.0	2022.1	1116.7	640.3	353.1	190.2
		S	5058.5	2512.3	1418.2	829.9	474.9	272.5
	30	B	3074.0	1416.4	713.7	350.0	148.6	34.6
		S	3976.8	1817.4	940.0	483.0	230.9	101.6
	60	B	1914.9	684.9	190.5	0	0	0
		S	2557.8	950.8	328.2	0	0	0
	90	B	755.8	0	0	0	0	0
		S	1230.2	95.0	0	0	0	0
100	0	B	4058.0	1900.3	1040.8	596.3	374.7	235.2
		S	5058.5	2392.5	1324.9	775.8	451.8	253.3
	30	B	3074.0	1343.5	650.6	306.6	118.8	16.3
		S	3976.8	1725.3	869.0	440.9	208.2	79.9
	60	B	1914.9	628.6	141.0	0	0	0
		S	2557.8	903.8	258.8	0	0	0
	90	B	755.8	0	0	0	0	0
		S	1230.2	82.2	0	0	0	0

* B→ Bituminous coal, S→ Subbituminous coal

Table 4.11: Surplus gas produced at different degrees of metallization for different post combustion ratio and heat transfer efficiency for bituminus and subbituminus coal when 40% O₂-enriched air is used for post combustion

Heat transfer efficiency	PCR		0	10	20	30	40	50
	% M							
70	0	B	4058.0	2284.3	1351.2	775.8	423.9	242.2
		S	5058.5	2882.0	1713.3	1007.9	562.5	337.1
	30	B	3074.0	1657.9	822.0	506.4	254.8	100.7
		S	3976.8	2148.8	1214.6	687.8	371.8	183.6
	60	B	1914.9	871.2	350.4	72.7	0	0
		S	2557.8	1197.5	560.5	212.2	0	0
	90	B	755.8	88.5	0	0	0	0
		S	1230.2	281.7	0	0	0	0
80	0	B	4058.0	2190.0	1269.9	736.8	403.8	225.5
		S	5058.5	2761.1	1616.3	948.4	540.9	317.9
	30	B	3074.0	1565.4	844.8	442.3	209.4	72.9
		S	3976.8	2004.6	1113.0	610.2	320.3	160.7
	60	B	1914.9	797.0	288.8	0	0	0
		S	2557.8	1102.7	456.8	0	0	0
	90	B	755.8	28.1	0	0	0	0
		S	1230.2	194.9	0	0	0	0
90	0	B	4058.0	2082.5	1189.8	690.0	384.5	209.2
		S	5058.5	2608.9	1507.1	897.8	515.9	296.1
	30	B	3074.0	1478.4	768.4	394.0	172.9	50.7
		S	3976.8	1904.0	1007.3	535.9	260.8	130.8
	60	B	1914.9	728.6	228.5	0	0	0
		S	2557.8	984.5	375.6	0	0	0
	90	B	755.8	0	0	0	0	0
		S	1230.2	142.5	0	0	0	0
100	0	B	4058.0	2024.4	1084.5	644.3	364.8	193.3
		S	5058.5	2377.0	1409.9	836.8	491.1	276.7
	30	B	3074.0	1403.2	702.1	341.4	141.8	30.7
		S	3976.8	1799.7	933.4	483.2	244.0	96.6
	60	B	1914.9	627.7	157.6	0	0	0
		S	2557.8	958.6	301.6	0	0	0
	90	B	755.8	0	0	0	0	0
		S	1230.2	102.9	0	0	0	0

* B→ Bituminus coal, S→ Subbituminus coal

Table 4.12: Surplus gas produced at different degrees of metallization for different post combustion ratio and heat transfer efficiency for bituminous and subbituminous coal when air is used for post combustion

Heat transfer efficiency	PCR		0	10	20	30	40	50
	% M							
70	0	B	4058.0	2410.3	1470.7	858.1	478.9	275.5
		S	5058.5	3034.1	1856.1	1117.7	631.1	376.9
	30	B	3074.0	1756.0	904.0	574.3	304.0	129.5
		S	3976.8	2243.0	1325.8	771.8	434.3	223.2
	60	B	1914.9	939.9	414.9	123.8	0	0
		S	2557.8	1275.0	640.6	284.9	0	0
	90	B	755.8	132.3	0	0	0	0
		S	1230.2	335.2	0	0	0	0
80	0	B	4058.0	2312.3	1380.0	817.8	455.9	254.8
		S	5058.5	2916.3	1756.1	1045.5	605.7	361.1
	30	B	3074.0	1661.9	929.7	506.3	251.2	100.7
		S	3976.8	2121.9	1222.8	689.5	374.4	193.2
	60	B	1914.9	863.2	384.5	0	0	0
		S	2557.8	1184.9	529.8	0	0	0
	90	B	755.8	70.9	0	0	0	0
		S	1230.2	253.3	0	0	0	0
90	0	B	4058.0	2204.7	1293.7	770.0	432.1	237.8
		S	5058.5	2736.2	1624.1	992.2	580.8	336.5
	30	B	3074.0	1567.2	849.6	450.7	209.2	79.3
		S	3976.8	2014.4	1104.0	614.7	306.3	164.2
	60	B	1914.9	792.7	283.5	0	0	0
		S	2557.8	1061.5	507.4	0	0	0
	90	B	755.8	0	0	0	0	0
		S	1230.2	191.2	0	0	0	0
100	0	B	4058.0	2145.7	1186.1	717.7	411.8	225.5
		S	5058.5	2507.1	1536.5	925.5	549.0	317.8
	30	B	3074.0	1488.9	774.8	393.5	176.7	52.6
		S	3976.8	1888.6	1020.5	548.6	294.5	124.0
	60	B	1914.9	689.5	214.6	0	0	0
		S	2557.8	1028.6	364.2	0	0	0
	90	B	755.8	0	0	0	0	0
		S	1230.2	147.9	0	0	0	0

* B→ Bituminous coal, S→ Subbituminous coal

Table 4.13: Export gas produced at different degrees of metallization for different post combustion ratio and heat transfer efficiency for bituminous and subbituminous coal when pure oxygen is used for post combustion

Heat transfer efficiency	PCR		0	10	20	30	40	50
	% M							
70	0	B	4212.4	2360.1	1424.7	882.7	544.3	376.5
		S	5212.9	2926.4	1765.8	1094.3	674.7	466.3
	30	B	3319.9	1836.9	1113.2	708.8	470.9	329.9
		S	4131.2	2277.9	1381.4	878.3	583.4	407.9
	60	B	2406.6	1313.6	799.8	534.9	0	0
		S	3049.5	1629.4	997.3	662.7	0	0
	90	B	1493.4	790.3	0	0	0	0
		S	1967.8	980.9	0	0	0	0
80	0	B	4212.4	2259.5	1347.9	838.7	525.9	360.7
		S	5212.9	2807.5	1674.9	1042.6	654.6	447.5
	30	B	3319.9	1745.7	1031.3	647.8	428.8	302.8
		S	4131.2	2169.1	1285.5	805.3	532.5	385.6
	60	B	2406.6	1242.1	738.2	0	0	0
		S	3049.5	1531.2	896.1	0	0	0
	90	B	1493.4	738.5	0	0	0	0
		S	1967.8	893.1	0	0	0	0
90	0	B	4212.4	2156.5	1271.1	794.7	507.5	344.6
		S	5212.9	2666.7	1572.6	984.3	629.3	426.9
	30	B	3319.9	1662.3	959.6	595.9	394.5	280.5
		S	4131.2	2063.3	1185.9	728.9	476.9	347.5
	60	B	2406.6	1176.6	682.2	0	0	0
		S	3049.5	1442.5	819.9	0	0	0
	90	B	1493.4	0	0	0	0	0
		S	1967.8	832.6	0	0	0	0
100	0	B	4212.4	2054.7	1195.2	750.7	489.1	329.6
		S	5212.9	2546.9	1479.3	930.2	606.2	407.7
	30	B	3319.9	1589.4	896.4	552.5	364.7	262.2
		S	4131.2	1971.2	1114.9	686.8	454.1	325.8
	60	B	2406.6	1120.3	632.7	0	0	0
		S	3049.5	1395.5	750.5	0	0	0
	90	B	1493.4	0	0	0	0	0
		S	1967.8	819.8	0	0	0	0

* B→ Bituminous coal, S→ Subbituminous coal

Table 4.14: Export gas produced at different degrees of metallization for different post combustion ratio and heat transfer efficiency for bituminous and subbituminous coal when 40% O₂ enriched air is used for post combustion

Heat transfer efficiency	PCR		0	10	20	30	40	50
	% M							
70	0	B	4212.4	2438.7	1505.6	930.2	578.3	396.6
		S	5212.9	3036.4	1867.7	1162.3	716.9	491.5
	30	B	3319.9	1903.8	1067.9	752.3	500.7	346.6
		S	4131.2	2370.7	1460.5	933.7	615.7	429.5
	60	B	2406.6	1362.9	842.1	564.4	0	0
		S	3049.5	1689.2	1052.2	703.9	0	0
	90	B	1493.4	826.1	0	0	0	0
		S	1967.8	1019.3	0	0	0	0
80	0	B	4212.4	2344.4	1424.3	891.2	558.2	379.9
		S	5212.9	2915.5	1770.7	1102.8	695.3	472.3
	30	B	3319.9	1811.3	1090.7	688.2	455.3	318.8
		S	4131.2	2250.5	1358.9	856.1	566.2	406.6
	60	B	2406.6	1288.7	780.5	0	0	0
		S	3049.5	1594.4	948.5	0	0	0
	90	B	1493.4	765.7	0	0	0	0
		S	1967.8	932.5	0	0	0	0
	0	B	4212.4	2236.9	1344.2	844.4	538.9	363.6
		S	5212.9	2763.3	1661.5	1052.2	670.3	450.5
90	30	B	3319.9	1724.3	1014.3	639.9	418.8	296.6
		S	4131.2	2149.9	1253.2	781.8	506.7	376.7
	60	B	2406.6	1220.3	720.2	0	0	0
		S	3049.5	1476.2	867.3	0	0	0
	90	B	1493.4	0	0	0	0	0
		S	1967.8	880.1	0	0	0	0
100	0	B	4212.4	2178.8	1238.9	798.7	519.2	347.7
		S	5212.9	2531.4	1564.3	991.2	645.5	431.1
	30	B	3319.9	1649.1	948.0	587.3	387.7	276.6
		S	4131.2	2045.6	1179.3	729.1	489.9	342.5
	60	B	2406.6	1119.4	649.3	0	0	0
		S	3049.5	1450.3	793.3	0	0	0
	90	B	1493.4	0	0	0	0	0
		S	1967.8	840.5	0	0	0	0

* B→ Bituminous coal, S→ Subbituminous coal

Table 4.15: Export gas produced at different degrees of metallization for different post combustion ratio and heat transfer efficiency for bituminous and subbituminous coal when air is used for post combustion

Heat transfer efficiency	PCR		0	10	20	30	40	50
	% M							
70	0	B	4212.4	2564.7	1625.1	1012.5	633.3	429.9
		S	5212.9	3188.5	2010.5	1266.1	785.5	531.3
	30	B	3319.9	2001.9	1150.1	820.2	549.9	375.4
		S	4131.2	2488.9	1571.7	1017.7	680.2	469.1
	60	B	2406.6	1431.6	906.6	615.5	0	0
		S	3049.5	1766.7	1132.3	776.6	0	0
	90	B	1493.4	869.9	0	0	0	0
		S	1967.8	1072.8	0	0	0	0
80	0	B	4212.4	2466.7	1534.4	972.2	610.3	409.2
		S	5212.9	3070.7	1910.5	1199.9	760.1	515.5
	30	B	3319.9	1907.8	1175.6	752.2	497.1	345.6
		S	4131.2	2367.8	1468.7	935.4	620.3	439.1
	60	B	2406.6	1354.9	840.2	0	0	0
		S	3049.5	1676.6	1021.5	0	0	0
	90	B	1493.4	808.5	0	0	0	0
		S	1967.8	990.9	0	0	0	0
90	0	B	4212.4	2359.1	1448.1	924.4	586.5	392.2
		S	5212.9	2890.6	1778.5	1146.6	735.2	490.9
	30	B	3319.9	1813.2	1095.5	696.6	455.1	325.2
		S	4131.2	2260.3	1349.9	860.6	552.2	410.1
	60	B	2406.6	1284.4	775.2	0	0	0
		S	3049.5	1553.2	999.1	0	0	0
	90	B	1493.4	0	0	0	0	0
		S	1967.8	928.8	0	0	0	0
100	0	B	4212.4	2300.1	1340.5	872.1	566.2	379.9
		S	5212.9	2661.5	1690.9	1079.9	703.4	472.2
	30	B	3319.9	1734.8	1020.7	639.4	422.6	298.5
		S	4131.2	2134.5	1216.4	794.5	540.4	369.9
	60	B	2406.6	2181.2	706.3	0	0	0
		S	3049.5	1520.3	855.9	0	0	0
	90	B	1493.4	0	0	0	0	0
		S	1967.8	885.5	0	0	0	0

* B→ Bituminous coal, S→ Subbituminous coal

Table 4.16 : Pure oxygen required for post combustion at different degrees of metallization for different post combustion ratio and heat transfer efficiency for bituminous and subbituminous coal

Heat transfer efficiency	PCR		0	10	20	30	40	50
	% M							
70	0	B	0	292.9	447.1	543.7	608.4	656.9
		S	0	333.9	507.4	617.4	690.5	744.3
	30	B	0	238.1	364.2	442.0	500.8	566.0
		S	0	272.1	416.2	501.9	567.1	641.4
	60	B	0	183.3	280.7	340.3	393.2	475.1
		S	0	209.1	319.4	385.3	444.6	537.3
	90	B	0	128.5	197.2	238.6	285.6	384.2
		S	0	147.5	212.5	268.7	321.3	433.3
80	0	B	0	280.3	419.9	504.2	562.0	609.3
		S	0	321.2	479.1	572.8	637.9	690.5
	30	B	0	227.5	338.2	404.5	453.6	503.5
		S	0	259.2	382.5	458.5	513.3	569.3
	60	B	0	174.7	256.5	304.8	345.2	397.7
		S	0	197.3	291.9	344.2	389.9	448.2
	90	B	0	122.9	174.8	205.1	236.8	291.9
		S	0	141.5	197.6	230.9	265.2	324.7
90	0	B	0	267.6	392.2	464.7	515.7	561.7
		S	0	305.3	448.5	527.0	584.2	635.7
	30	B	0	216.3	316.1	373.5	414.3	457.6
		S	0	246.9	362.5	423.0	468.7	516.8
	60	B	0	165.1	240.1	282.3	326.3	352.3
		S	0	190.3	275.3	318.9	368.1	396.7
	90	B	0	118.2	163.9	193.2	213.1	249.2
		S	0	134.9	187.9	217.2	238.9	279.3
100	0	B	0	255.0	364.5	425.2	469.4	514.1
		S	0	295.3	418.9	481.3	531.6	581.3
	30	B	0	207.5	295.9	345.6	381.1	418.0
		S	0	238.1	339.2	396.6	436.7	478.2
	60	B	0	160.5	227.9	266.0	292.8	321.9
		S	0	182.9	338.1	300.7	329.3	362.3
	90	B	0	112.5	157.9	186.4	204.5	225.8
		S	0	128.1	179.4	205.6	228.7	250.4

* B→ Bituminous coal, S→ Subbituminous coal

Table 4.17 : 40% O₂ enriched air for post combustion at different degrees of metallization for different post combustion ratio and heat transfer efficiency for bituminous and subbituminous coal

Heat transfer efficiency	PCR		0	10	20	30	40	50
	% M							
70	0	B	0	764.2	1193.4	1470.0	1661.3	1806.8
		S	0	1065.8	1658.2	2041.3	2307.1	2490.9
	30	B	0	620.9	970.2	1195.1	1351.6	1471.2
		S	0	864.9	1348.2	1652.8	1874.6	2052.1
	60	B	0	477.5	747.1	920.2	1041.9	1135.7
		S	0	665.5	1021.5	1281.5	1447.5	1573.3
	90	B	0	334.2	523.8	645.3	732.3	800.1
		S	0	465.9	729.9	898.5	1020.6	1116.5
80	0	B	0	730.1	1115.5	1356.3	1520.6	1643.6
		S	0	1018.4	1515.9	1861.3	2130.5	2241.5
	30	B	0	592.9	906.9	1102.5	1236.3	1336.9
		S	0	825.9	1261.3	1527.3	1723.9	1843.7
	60	B	0	455.8	698.3	848.6	951.9	1030.3
		S	0	634.7	745.4	1177.9	1320.6	1426.9
	90	B	0	318.7	489.7	594.8	667.6	723.6
		S	0	443.6	647.5	827.6	930.5	1008.6
	0	B	0	696.2	1037.5	1242.8	1380.0	1480.3
		S	0	970.1	1441.3	1712.6	1915.4	2045.6
90	30	B	0	482.5	843.6	1009.9	1120.9	1202.6
		S	0	673.4	1171.9	1397.6	1552.4	1666.8
	60	B	0	392.9	649.6	777.0	861.9	924.3
		S	0	547.3	883.3	1081.5	1200.8	1288.9
	90	B	0	303.3	455.7	544.2	602.9	647.2
		S	0	422.7	634.5	754.3	838.7	902.6
	0	B	0	661.9	959.6	1129.7	1239.4	1317.1
		S	0	922.9	1311.5	1554.9	1708.9	1807.2
	30	B	0	537.2	780.3	917.5	1005.7	1068.3
		S	0	749.2	1098.5	1269.2	1410.6	1469.9
100	60	B	0	412.5	601.5	705.6	772.1	819.5
		S	0	574.8	834.1	976.5	1066.7	1132.5
	90	B	0	287.8	421.7	493.7	538.3	570.7
		S	0	400.1	581.9	686.7	748.3	795.2

* B→ Bituminous coal, S→ Subbituminous coal

Table 4.18 : Air required for post combustion at different degrees of metallization for different post combustion ratio and heat transfer efficiency for bituminous and subbituminous coal

Heat transfer efficiency	PCR		0	10	20	30	40	50
	% M							
70	0	B	0	1557.1	2526.8	3193.4	3682.2	4053.5
		S	0	2172.1	3492.1	4294.7	4936.7	5439.9
	30	B	0	1265.1	2054.4	2596.6	2995.3	3285.5
		S	0	1734.3	2825.9	3611.4	4078.5	4552.7
	60	B	0	973.2	1581.9	1999.7	2308.3	2537.5
		S	0	1338.2	2210.5	2778.6	3200.5	3519.7
	90	B	0	681.2	1109.5	1402.9	1621.4	1789.5
		S	0	939.2	1523.6	1945.8	2251.9	2485.6
80	0	B	0	1483.5	2348.5	2923.1	3333.7	3643.4
		S	0	2062.5	3260.5	4037.6	4499.5	4951.7
	30	B	0	1205.0	1908.4	2375.3	2710.4	2962.7
		S	0	1681.2	2651.3	3301.1	3690.7	4231.7
	60	B	0	926.6	1469.1	1827.4	2087.0	2282.4
		S	0	1293.5	2021.5	2528.7	2901.5	3173.5
	90	B	0	648.1	1029.3	1279.5	1463.7	1601.9
		S	0	900.5	1421.9	1784.6	2040.3	2233.4
90	0	B	0	1409.2	2170.2	2652.9	2985.3	3233.2
		S	0	1961.3	3001.5	3639.6	4064.3	4460.3
	30	B	0	1144.4	1763.3	2154.1	2425.6	2627.1
		S	0	1537.5	2455.3	2891.3	3312.5	3634.7
	60	B	0	879.7	1356.2	1654.9	1864.9	2020.7
		S	0	1220.5	1891.6	2306.1	2600.3	2807.9
	90	B	0	614.9	949.2	1156.0	1306.3	1414.4
		S	0	850.9	1323.5	1602.5	1793.6	1972.5
100	0	B	0	1335.1	1991.9	2382.6	2636.8	2823.1
		S	0	1858.5	2711.5	3312.7	3578.5	3838.1
	30	B	0	1084.0	1617.6	1932.4	2140.7	2291.0
		S	0	1498.5	2241.9	2655.2	2954.9	3165.3
	60	B	0	832.9	1243.3	1482.5	1644.2	1758.9
		S	0	1160.7	1721.9	2068.3	2291.4	2432.3
	90	B	0	581.8	869.1	1032.6	1148.3	1226.8
		S	0	808.9	1211.9	1435.7	1601.5	1702.5

* B→ Bituminous coal, S→ Subbituminous coal

Table 4.19 : Volume of gas in the melter gasifier at various post combustion ratio for different post combustion medium

PCM	% PCR %M	0	10	20	30	40	50
Oxygen	0	4291.2	2810.3	2094.1	1666.8	1386.7	1180.6
	30	3478.5	2282.9	1701.7	1355.7	1127.4	366.3
	60	2673.3	1754.8	1309.3	1044.6	868.1	743.9
	90	1868.2	1227.1	916.9	730.6	608.7	523.7
40% O ₂ - enriched air	0	4291.2	3143.4	2554.9	2191.9	1950.5	1778.3
	30	3478.5	2554.2	2077.1	1783.6	1586.4	1446.3
	60	2673.3	1964.9	1594.6	1371.3	1222.3	1114.2
	90	1868.2	1370.1	1116.7	962.9	856.2	782.2
Air	0	4291.2	3702.8	3363.9	3151.8	3004.9	2894.7
	30	3478.5	3004.6	2735.9	2561.2	2440.8	2355.2
	60	2673.3	2313.0	2101.8	1970.5	1881.8	1815.7
	90	1868.2	1614.8	1473.8	1379.8	1317.6	1276.3

* PCM : Post Combustion Medium

Table 4.20: Gas compositions in the melter-gasifier at various post combustion ratio for various post combustion medium

PCM	%PCR	%CO	%CO ₂	%H ₂	%H ₂ O	%N ₂
Oxygen	0	73.6	0	25.8	0	0.59
	10	66.1	7.34	23.4	2.6	0.59
	20	58.6	14.6	20.9	5.2	0.59
	30	51.1	21.9	18.4	7.9	0.6
	40	43.7	29.1	15.9	10.6	0.6
	50	36.3	36.3	13.4	13.4	0.6
40% O ₂ - enriched Air	0	73.6	0	25.8	0	0.59
	10	61.5	6.8	21.8	2.4	7.5
	20	51.0	12.8	18.2	4.5	13.5
	30	41.8	17.9	15.0	6.4	18.8
	40	33.7	22.5	12.2	8.1	23.4
	50	26.5	26.5	9.7	9.7	27.6
Air	0	73.6	0	25.8	0	0.59
	10	55.7	6.2	19.7	2.2	16.3
	20	42.7	10.7	15.2	3.8	27.7
	30	32.8	14.1	11.7	5.0	36.3
	40	25.1	16.1	9.0	6.0	43.1
	50	18.9	18.9	6.8	6.8	48.6

* PCM : Post Combustion Medium

Table 4.21: Gas temperature at different post combustion ratio for different heat transfer efficiency and post combustion medium

Post combustion medium	HTE PCR	0	20	40	60	80	100
Pure Oxygen	20	2541.1	2301.8	2057.7	1804.1	1540.9	1273.0
	30	3018.4	2686.4	2348.6	1999.5	1639.2	1273.0
	40	3373.1	2974.9	2569.3	2149.2	1714.8	1273.0
	50	3691.5	3236.1	2771.0	2287.4	1784.9	1273.0
40% O ₂ - enriched air	20	2306.4	2104.7	1903.0	1696.4	1484.7	1273.0
	30	2641.5	3351.8	2756.8	2156.7	1550.6	1273.0
	40	2905.3	2582.8	2262.4	1930.6	1579.3	1273.0
	50	3118.1	2755.3	2395.1	2015.2	1590.0	1273.0
Pure air	20	2033.3	1886.3	1739.2	1587.1	1430.1	1273.0
	30	2237.4	2049.2	1861.4	1668.6	1470.8	1273.0
	40	2386.5	2178.8	1966.1	1728.4	1500.7	1273.0
	50	2500.2	2264.7	2024.3	1778.8	1523.4	1273.0
700°K Preheated air	20	2139.1	1975.8	1816.4	1639.1	1446.2	1273.0
	30	2369.6	2160.3	1956.5	1736.2	1502.3	1273.0
	40	2537.4	2294.3	2051.7	1798.5	1538.4	1273.0
	50	2665.1	2395.4	2129.3	1849.8	1561.5	1273.0
1000° K Preheated air	20	2221.0	2043.9	1862.6	1672.9	1475.1	1273.0
	30	2471.0	2245.1	2014.5	1774.9	1526.2	1273.0
	40	2653.1	2391.2	2124.6	1848.6	1563.2	1273.0
	50	2792.3	2502.7	2208.5	1904.7	1591.3	1273.0

Table 4.22: Degree of dissociation of CO₂ at various post combustion ratio for different post combustion medium

PCR \ PCM	10	20	30	40	50
Oxygen, 900°K	0	0	0	21.0	40.0
Oxygen, 600°K	0	0	0	14.8	33.9
Oxygen, 298°K	0	0	0	7.96	28.9
Preheated air, 1000°K	0	0	0	0	0
Preheated air, 700°K	0	0	0	0	0
40% O ₂ -enriched air, 298°K	0	0	0	0	0
Air, 298°K	0	0	0	0	0

* PCR: Post combustion ratio

* PCM: Post combustion medium

Table 4.23: Optimum metallization at various post combustion ratio

%PCR	Optimum Metallization (%)
10	87
20	69
30	54
40	42
50	32